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ACTIVATED CARBON TREATMENT OF RAW SEWAGE IN SOLIDS-CONTACT CLARIFIERS

Richard L. Beebe

Westinghouse Electric Corporation

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ACTIVATED CARBON TREATMENT OF RAW SEWAGE IN SOLIDS-CONTACT CLARIFIERS

by

Richard L. Beebe

Project No. 17050 EGI
Contract No. 14-12-586

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The maximum carbon adsorptive capacity for filtrable COD was 0.50 to 0.55 g COD/g carbon. This capacity was achieved whenever the concentrations of influent COD and carbon matched or exceeded that ratio (adsorptive-resistant COD excluded). Carbon requirements were 55 to 60 percent of theoretical two-stage countercurrent adsorption system requirements.

Assuming regeneration cycles 85 percent of the carbon feed, respective treatment cost estimates for 10-mgd and 100-mgd plants were 13.9¢ and 11.2¢ per thousand gallons. (EPA)

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ABSTRACT

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This report was submitted in fulfillment of Contract No. 14-12-586, Project No. 17050 EGI, between the Environmental Protection Agency and INFILCO Division, Westinghouse Electric Corporation.

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CONCLUSIONS

Degritted raw municipal sewage can be treated efficiently and economically with powdered activated carbon and coagulants in a continuous high-rate process.

The process effluent is equal to or better in quality than that produced in a parallel activated sludge plant.

Activated carbon utilization is maximized by contacting the carbon with the waste stream in two series-connected high-rate recirculating-slurry solids-contact clarifiers. Each clarifier is capable of exceeding single-stage carbon adsorption efficiency because the carbon flows countercurrent to the waste stream during a portion of its circulation. Fresh carbon is fed to the recirculation zone of the downstream clarifier and is then transferred countercurrent to the waste stream between the clarifiers. This equipment and flow arrangement provides an effective two-stage countercurrent contacting system.

Under actual test conditions, the powdered activated carbon requirement was conservatively calculated to be not more than 20 to 25 percent of single-stage batch contacting or 55 to 60 percent of two-stage countercurrent adsorption requirements.

The unique compound-contacting system makes possible nearly complete removal of carbon-adsorbable COD-producing compounds at peak carbon loadings. The test sewage stream contained a significant amount of adsorption-resistant compounds which limited the removal of filtrable TOC and COD to 68.0 and 69.9 percent, respectively, of that in the feed stream. The activated carbon adsorption capacity for filtrable COD was between 0.50 and 0.55 g COD per g of carbon. This loading range was achieved whenever the adsorbable COD and carbon were present at this or higher ratios. The carbon adsorption capacity for filtrable TOC was about 0.17 g per g of carbon.

The optimum carbon dosage can be calculated from the concentration of available adsorbate and the carbon adsorption capacity. In this study, this was found to be 132 percent of the feed stream filtrable COD or about 1 lb/1000 gal. Higher carbon dosages removed very little additional adsorbate resulting in reduced adsorbate loading per unit weight. Lower carbon dosages resulted in capacity carbon loading, but reduced the degree of treatment.

The process removes the major portion of the total TOC and COD from the sewage by coagulation and sedimentation of the raw sewage suspended solids. The powdered activated carbon and its adsorption process are combined with and proceed simultaneously with the solids removal. The combined systems, along with effluent filtration resulted in mean total-TOC and -COD removals of 88.1 and 88.7 percent as compared with the activated sludge plant removals of 86.6 and 88.2 percent, respec-

tively. BOD removals were 90 percent for the carbon adsorption system and 88 percent for the activated sludge system.

Laboratory studies are particularly useful in determining the more effective and economical treating chemicals. This study determined that Aqua Nuchar A and Darco S-51 powdered activated carbons were equally effective adsorbents for this sewage and they were effectively coagulated, in conjunction with raw sewage solids by 6 to 8 mg/l of C-31 polyelectrolyte and 10 to 15 mg/l of filter alum. Operation of a 28,000-gpd pilot plant substantially confirmed these findings.

Unit chemical costs are relatively high. Aqua Nuchar A is the more economical product and was used during most of the pilot plant tests. Darco S-51 was estimated to cost 0.6¢/1000 gal. more than Aqua Nuchar A but its physical characteristics, i.e., higher bulk density and greater particle size, offer advantages in floc density, floc dewatering, housekeeping and reduced storage volume requirements.

Neither C-31 polyelectrolyte nor filter alum, when used singly, was suitable for the needs of the process. Without filter alum the required polymer dosage was undesirably costly. Alum in quantities needed for good clarification produced excessive volumes of sludge. When combined, the polymer was observed to coagulate the carbon effectively while the alum was instrumental in controlling turbidity causing materials present in the sewage.

During the pilot plant study the limited pumping capacity permitted a maximum ACCELATOR® clarifier overflow rate of 1.5 gpm/ft^2. However, subsidence tests on the solids slurry taken from the draft tube of the ACCELATOR mechanism indicated that overflow rates equivalent to or greater than 2.25 gpm/ft^2 were easily achievable. Therefore, based upon these subsidence tests and other in-house experience with ACCELATOR units, a conservative overflow rate of 2.25 gpm/ft^2 was used as a design parameter to establish the cost of 10- and 100-mgd treatment facilities. At this overflow rate the experimental system, including final gravity filtration, would have a retention time of 63 minutes.

As discharged from the experimental system the solids content of the blowdown mixture of raw sewage solids and carbon was 2.5 percent. External thickening would be required for economical carbon regeneration. Regeneration and reuse of the spent carbon were not studied.

Filtration of the treated stream is practical. Although it may not be required for some potential applications of the process, the existence of the polishing filter permits some reduction in coagulant dose by confining the resulting increase in carbon-floc carryover to the plant site. The filter assures carbon-free effluent when necessary. No carbon passed through the pilot plant filter except at times of experimental extremes. The filter was operated at surface loadings up to 2.8 gpm/ft^2 with a maximum headloss limit of 4.5 ft. Nominal

wash water consumption was 3.0 percent of that filtered during filter runs of 25 hours, although these figures varied widely with the experimental program. Modest biological growth in the underdrain system was controlled but not eliminated by chlorinated water backwashing.

Despite varied operating conditions, the system exhibits remarkably uniform adsorption process performance. The clarifier carbon inventory was capable of maintaining effluent quality even with up to 8 hours loss of carbon feed.

Coagulant consumption for the complete process is double the average single treatment dosage previously stated because there are two clarification steps. On a 10-mgd scale, assuming carbon, polyelectrolyte and alum consumption of 118, 14 and 25 mg/l, respectively, the total estimated cost of treatment by the pilot plant process is 13.9¢/1000 gal. A 100-mgd plant of similar design would produce treated water at a cost of 11.2¢/1000 gal. Both cost figures incorporate a 1.9¢/lb rate for regeneration of spent carbon slurry assuming the raw sewage solids contained in the slurry present no special cost problems. Both plants feature 85 percent carbon recovery with 15 percent new carbon makeup. Solid and liquid waste handling and disposal problems are minimal for either plant.

RECOMMENDATIONS

This pilot plant investigation demonstrated the feasibility of treating raw sewage in a short-retention-time carbon-adsorption/flocculation system using new activated carbon. Process economics depend upon recovery and reuse of regenerated carbon from the mixture of raw sewage solids and spent carbon in the blowdown slurry. Effort should be directed at determining if such regeneration is possible. Means of concentrating the blowdown slurry solids to optimum concentrations for economical regeneration should be explored simultaneously. Work to date has generally been restricted to thickening and regenerating spent carbon from tertiary treatment processes.

Adsorption-resistant organic materials in the process effluent will vary in concentration and character between plants. Methods for controlling or removing this fraction should be investigated, possibly with present-day processes for nutrient removal which may be desirable or necessary.

No biological growth inhibiters were used in the carbon contacting portion of the pilot system during this study. Potential process interferences as the result of biological growth were not encountered; however, this may not be typical for all applications. The effectiveness and cost of different means of controlling biological growth throughout the system warrant attention.

The presence of filtrable solids in the sewage being processed interfered with making a precise theoretical evaluation of the pilot plant adsorption capabilities in terms of accepted laboratory control procedures. Basic process efficiency should be determined and equated to known adsorption theory by operating a reduced scale plant on a laboratory prepared feed stream of adsorbate, monitored with suitable adsorption isotherm studies. The relationships established would simplify the problems of predicting performance for suggested practical applications.

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INTRODUCTION

Background

Adsorption is an established process for separating relatively small quantities of organic impurities or valuable products from dilute solutions. Activated carbon is one of the more efficient and widely used adsorbents.

Activated carbon adsorption has been used in water purification for more than 50 years.[1] It has successfully competed with other purification or recovery processes in industry, and more recently, has been applied to advanced wastewater treatment.[2] The need to control the amount and types of waste products fouling the environment has been the driving force for advanced wastewater treatment research. A secondary impetus is the benefit derived from recovering wastewater for reuse, thus preserving existing supplies of high quality raw waters.

The amount of carbon required to improve the quality of sewage plant effluents to acceptable levels may be very high because adsorption is controlled by aspects of kinetics and adsorption equilibrium.[3] A traditional single-stage powdered activated carbon contacting system is relatively inefficient and costly to operate because of the equilibrium adsorption phenomena. Continuous multistage countercurrent systems offer improved carbon economy at the expense of greater capital costs for the more extensive equipment system. Cost factors have favored the use of granular activated carbon in columnar systems with regeneration and reuse of the carbon.[4] Recently compound systems of contacting powdered carbon have been tested that operate with improved economy, approaching the cost of granular carbon systems provided regenerated carbon is used.

Two studies[5,6] utilized powdered activated carbon in coagulated slurry form recirculating in ACCELATOR clarifiers which are high-rate solids-contact treating units. These early investigations were confined to tertiary applications with the inherent disadvantage of working with low potential adsorption equilibrium conditions. With a low initial concentration of adsorbate, the carbon utilization was limited. Both studies, however, produced carbon utilization levels equal or superior to theoretical systems based upon isotherms derived from laboratory jar tests.

Providing a higher concentration of adsorbate along with refinements in process design would substantially improve the amount of impurities adsorbed upon each measure of activated carbon. The highest concentration of adsorbate of interest occurs in the raw sewage entering the treatment plant and remains little changed if not unchanged as the sewage passes thru the primary clarifier. While substantial amounts of suspended matter are removed by screening, grit removal and primary

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sedimentation the soluble material flows on to the biological or chemical process units.

If a compound-contacting powdered activated carbon system could be operated on primary clarifier effluent or even screened and degritted raw sewage, more efficient use could be made of the carbon's adsorptive properties. A substantial amount of the operating cost would be offset by eliminating the need for traditional secondary treatment facilities and perhaps primary clarification and sludge disposal also. Certainly coarse screening (or comminution) and grit removal would be retained to protect equipment. A compact chemical-carbon slurry treating plant would be free of the biological problems associated with traditional secondary treatment facilities.

Process Description

A two-contactor countercurrent system was tested. Application of the process involved series operation of two solids-contact clarifier units of a type used widely for water treatment. New carbon was fed to the second unit only. The carbon feed for the first unit was the carbon slurry wasted from the second unit. Spent carbon was withdrawn from the system by blowdown from the first unit. The carbon treated and clarified liquid was filtered through a dual media gravity filter prior to discharge.

Objectives of This Investigation

The primary objective of the present investigation was a performance and cost evaluation of such a system for treatment of screened and degritted raw municipal sewage. Provisions made for processing primary clarifier effluent, if the treatment of screened and degritted raw sewage proved impractical, were not used.

Experimental Approach

A commercially available powdered activated carbon and flocculation agents to improve slurry settleability were selected on the basis of laboratory study.

These materials were then used during 27 consecutive pilot plant operating periods under various conditions. These operating periods have been grouped for reporting into seven phases, each grouping consisting of operating periods of related interest.

A high proportion of total pilot plant operating time was expended in investigating coagulant cost reduction possibilities.

First-phase operating periods established physical and hydraulic capabilities of the system and determined the treatability of the screened

and degritted raw sewage. Second-phase periods monitored the effects of reducing and combining coagulant feeds. The third phase demonstrated performance of a different powdered activated carbon followed by the fourth phase, a reference operating period. Fifth-phase operating periods further pursued the coagulant cost reduction objective using greater deviations in feed rates than previously used. Sixth-phase periods were intended to establish operating parameters using only an inorganic chemical coagulant. Seventh-phase periods compared results achieved with different dosages of powdered activated carbon.

During all operating periods, data were collected from the pilot plant and the adjacent municipal sewage treatment plant. Single-stage adsorption isotherms were developed at intervals for the feed stream and for the upstream clarifier effluent. Precise comparison of these data by application of adsorption theory[7] was not possible. Essential laboratory procedures introduced an irreconcilable datum shift between pilot plant results and isotherm results.

Peripheral laboratory analyses of 24-hour composite samples plus auxiliary grab samples were used to disclose operational parameters other than those directly associated with the adsorption process. These data, together with the pilot plant control records, reveal important practical operating requirements and results in other areas of interest.

Analytical Procedures

With certain exceptions, methods used in this study conformed to those in "Standard Methods."[8]

A modification to "Standard Methods" developed especially for low-level COD values was used. The complete procedure appears in the Appendix. Difficulty in obtaining reproducible results in IC (inorganic carbon) determinations with the total organic carbon analyzer coupled with the presence of 65 to 90 mg/l of IC in all samples prompted the abandonment of IC determinations. The deviation in IC determinations per sample approached the level of TOC to be measured. Inorganic carbon fractions of all samples for TOC determinations were removed by acidifying a 200-ml sample with 1 ml HCl and degassing with caustic-solution-scrubbed and filtered compressed air for 15 minutes. Replicate 20-microliter prepared samples were then manually injected into the TC (total carbon) combustion chamber of the furnace module. Laboratory trials revealed that this procedure reduced the indicated inorganic carbon residual to 1.0 mg/l C (the same as N_2-scrubbed samples) near the zero end of the analytical range (0-100 mg/l C) and to 1.2 mg/l C at the 35-mg/l TOC level. Filtered sample TOC determinations usually were less than 35 mg/l. Instrument calibration curves prepared in accordance with the manufacturer's directions were compensated to eliminate 1 mg/l of inorganic carbon residual.

Sets of samples prepared and analyzed for TOC in the laboratory were submitted to the EPA Advanced Waste Treatment Research Laboratory in Cincinnati, Ohio for comparison of results. The comparison determinations differed by 1 mg/l or less on a majority of the samples. Various operational difficulties with the several components of the TOC analyzer were detected and correct operation verified by frequent standardization. Standardization always preceded sample analyses.

All samples for which filtrable (filtrate) COD and TOC are reported were vacuum filtered through 0.45-micron Millipore filter discs prior to analysis. The filtrates were observed at times to contain discernable amounts of filtrable solids. Filtration was necessary in the case of powdered activated carbon treated samples as residual carbon particles contribute to the COD and TOC values. In order to provide a usable relationship between carbon treated samples and untreated samples, most of the untreated samples were filtered and reported in a like manner. Recently, vacuum filtering and also acidification and degassing procedures are recognized as altering the COD and TOC test results.[9] Volatile organics are lost in both procedures and this was verified during this study. A detailed investigation was not instituted and corrective procedures and equipment were not added to the program. The resulting error is not considered to alter the study conclusions significantly and permits direct data comparison with other studies using similar techniques.

Suspended solids determinations on the composite samples of screened and degritted raw sewage incorporated a preliminary coagulation step to facilitate filtering. Filter alum was used to treat the sample aliquot and a tap water blank. Sample results were corrected by the amount of the suspended solids produced in the blank. Direct vacuum filtration of the screened and degritted raw sewage aliquot was abandoned after encountering filter mat clogging with aliquots as small as 25 ml.

Certain ACCELATOR clarifier control tests performed at the pilot plant site are discussed hereinafter.

PREPARATORY LABORATORY INVESTIGATION

Objectives

An adsorbent-coagulant system was to be developed which would be physically manageable in the pilot plant equipment as well as produce process data. The initial feed stream was to be screened and degritted raw domestic sewage. The characteristics of the feed stream had to be determined. The adsorbent, activated carbon in powdered form, is available from different sources. It is manufactured from various raw materials by a number of different processes and may be specialized in its application.[10] Activated carbons were tested with aliquots of feed stream to assure selection of a material that would perform within the time limitations and hydraulic conditions imposed by the pilot plant. A flocculation system using this carbon in the feed stream had to be found. Inorganic coagulating agents were to be avoided if possible as they could be cumulative in the carbon when reactivation and reuse are concerned. The flocculated slurry must settle rapidly, leaving little if any residual carbon particles in the supernatant; and it would be advantageous if the slurry dewatered well on standing.

For the purposes of this study, the filtrable TOC of the feed is the primary adsorbate concentration indicator. Secondary indication was accomplished with filtrable COD determinations. Some impurities present may be adsorbed which do not contain appreciable TOC or exhibit COD while others with these characteristics may not be readily adsorbed. Sewage contains a varying combination of these materials, many of them not identified; so the TOC and COD determinations are a useful but incomplete pollution indicator.

Screened and Degritted Raw Sewage Quality

The City of Tucson, Arizona operates a 12.9 mgd step aeration activated sludge sewage treatment plant constructed in 1968. Raw sewage is passed through a mechanically-cleaned coarse bar screen and a grit-removal system prior to entering a flow-dividing weir box situated between the two primary clarifiers. Influent enters the bottom of the box, flows vertically upward and discharges to each of the clarifiers over two adjustable weir gates. Neither sediment nor floatable material accumulates in the space between weirs. This space was selected as the take-off point that would best provide a continuous representative supply to the pilot plant.

The sewage plant flow varies around design capacity at which the retention times are 2.0 hours in the primary clarifiers, 4.2 hours in the aeration basins and 2.1 hours in the final clarifiers. Physical and chemical characteristics of several raw sewage grab samples collected at random from the flow-dividing weir box are presented in Table 1.

TABLE 1

Screened and Degritted Raw Sewage Characteristics, mg/l
Tucson Sewage Treatment Plant

Date	11/12/69	11/14/69	11/18/69	12/3/69	1/14/70
Time	8:30 A.M.	8:30 A.M.	4:00 P.M.	7:00 A.M.	10:30 A.M.
Calcium (as Ca)	64	59	60	55	
Magnesium (as Mg)	14.6	14	16.5	17.0	
Sodium (as Na)	132	119	133	140	
Potassium (as K)	10.1	8.6	10.9	13.0	
Phosphorus (as P)	3.5	5.7	7.1	4.4	
Ammonia Nitrogen (as N)	26.7	29.5	26.8	19.5	
Silica (as SiO_2)	25.2	34	32	42	
Non-Carbonate Solids (as $CaCO_3$)	320	262	298	266	
Total Cations (as $CaCO_3$)	600	546	618	588	
Alkalinity (as $CaCO_3$)	280	298	306	304	
Hardness (as $CaCO_3$)	220	198	218	203	
Color (SU)	22	18	28	32	
Turbidity (JTU)	115	115	250	130	
Suspended Solids	202	206	374	165	
TOC (unfiltered)	-	124	169	85.0	180
TOC	33.0	25.0	61.0	23.0	47.5
COD (unfiltered)	-	-	-	-	740
COD	92	88	124	106	155
BOD_5 (unfiltered)	126	134	242	162	274
pH	8.1	7.9	7.9	7.9	-

Samples filtered thru 0.45-micron filter except as noted

Adsorption Isotherm Study

Two powdered activated carbon products were selected for evaluation against project requirements. These were Darco S-51, a product of Atlas Chemical Industries, Inc. and Aqua Nuchar A, a product of the Westvaco Corporation. A third product, Hydrodarco B, also manufactured by Atlas Chemical Industries was investigated briefly. It is marketed on a restricted basis.[11]

Darco S-51 was used in our earlier experiments and Aqua Nuchar A was utilized in an EPA pilot plant at Lebanon, Ohio. The rate of adsorbate uptake for both of these products had been found acceptable in the previous studies with secondary effluent. Adsorption efficiency, however, would need reevaluation with raw sewage.

Sufficient raw sewage was prefiltered through 0.45-micron filter discs to perform each carbon comparison study. A stock solution of each grade of carbon was prepared with special low-COD distilled water (appendix). The activated carbons were added to sets of five 250-ml aliquots of the prepared sewage at concentrations of 40, 160, 320, 800 and 1600 mg/l. A sixth aliquot in each set was left untreated as a control. The jars were stirred at 100 rpm for 60 minutes. Their contents were then filtered thru 0.45-micron filter discs, and filtrate TOC and/or COD determined.

Previous studies had shown the feasibility of handling the carbons as stock suspensions. To avoid performance differentials resulting from differences in wetting characteristics the suspensions were prepared not less than 24 hours prior to use and subjected to frequent manual shaking during use. The stirring speed of 100 rpm and the 60-minute stirring time also resulted from the previous experiments.

Freundlich curves of the data (Figures 1 and 2) were developed. The empirical Freundlich equation is generally written: $X/M = KC_t^{1/n}$, where X is the mg/l of adsorbate taken up by M mg/l of adsorbent applied. C_t is the mg/l of adsorbate remaining unadsorbed at equilibrium and K and 1/n are constants specific for the materials and conditions of test. When X/M is plotted against C_t on logarithmic paper, a straight line is commonly obtained which has a slope of 1/n and an intercept K at $C_t = 1$.

Slight variations in the original concentrations (C_o) of TOC and COD, perhaps caused by the passage of time, between the individual carbon test sets were retained in plotting the Freundlich curves. However, for positioning the curves of constant carbon dosage, an average C_o value was utilized.

With rare exception, the data from 19 separate isotherm experiments failed to plot as a straight line. This was true for all three carbons studied and for TOC- and COD-generated curves. Curves from TOC and COD

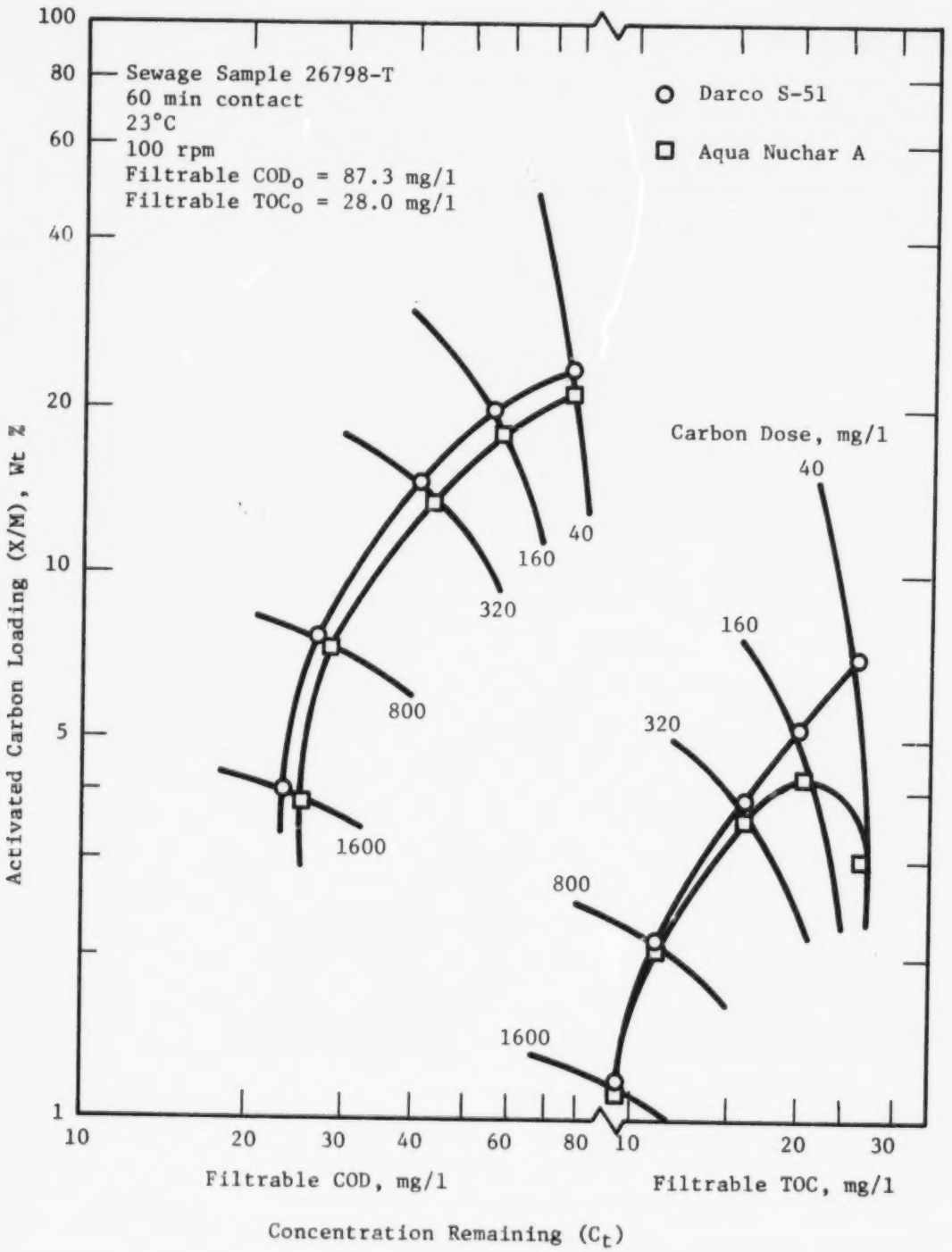

FIGURE 1: Adsorption Isotherm Carbon Comparison, Example A

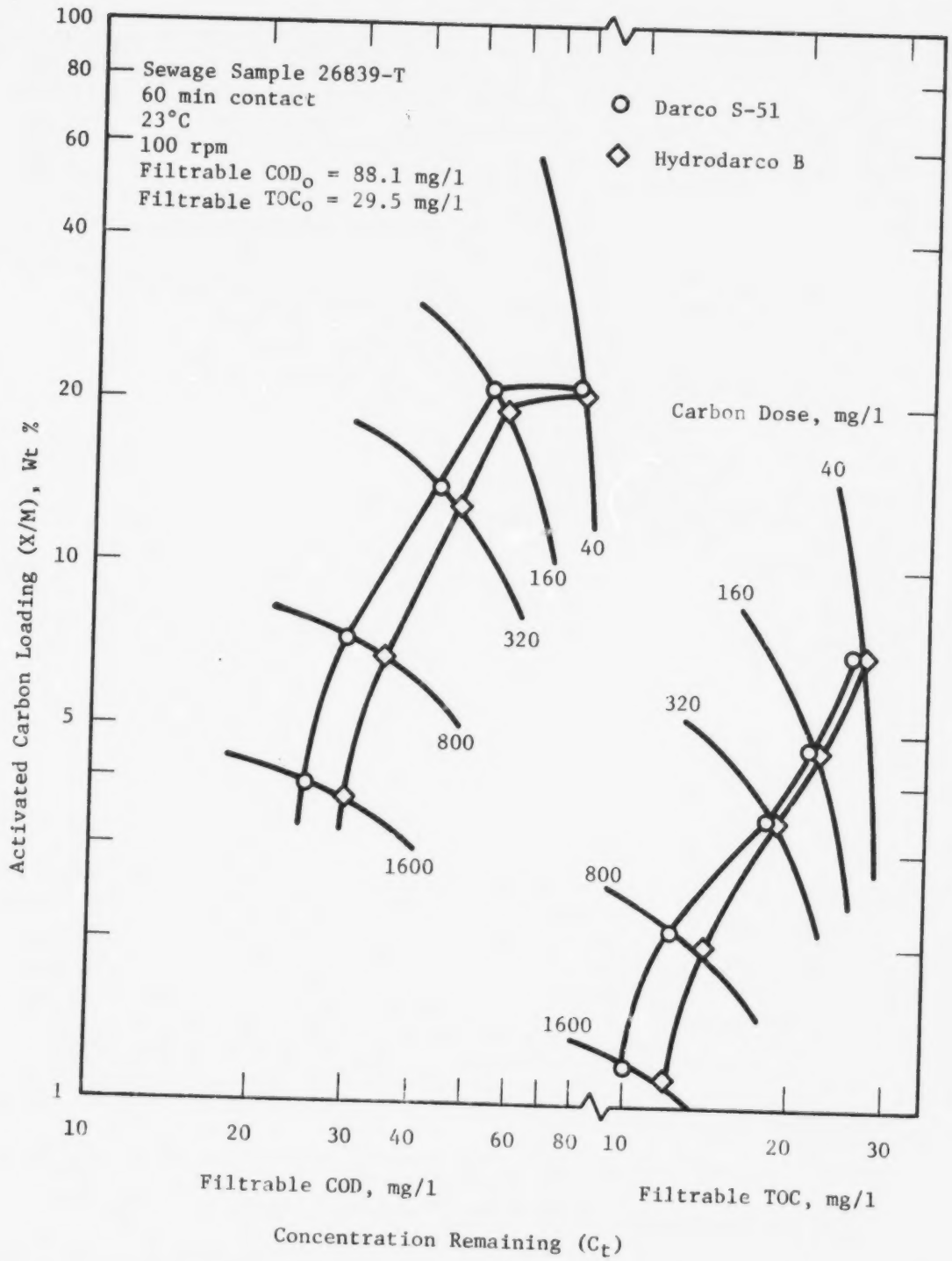

FIGURE 2: Adsorption Isotherm Carbon Comparison, Example B

data from an individual jar set differed in configuration as the ratio of COD to TOC deviated with carbon concentration. Utmost attention to technique did not resolve these questions. Each sewage sample produced a differently located and shaped isotherm. Figures 1 and 2 are representative of these variations. In general, a single-stage carbon loading limit was implied. At middle-range loadings, a relatively straight curve sometimes occurred. At lower loadings there was consistent evidence that residual adsorbate resisted adsorption regardless of the carbon under test. Figure 3, an extended-range isotherm illustrates these phenomena. There were small differences in the performances of the carbons studied when all isotherms are considered collectively. Darco S-51 usually exhibited a slightly higher capacity than Aqua Nuchar A which was slightly superior to Hydrodarco B. Selection of Aqua Nuchar A for the major portion of the program was finally made as a result of price differential.

The manufacturer states Nuchar active carbons are made from a residual organic material which is recovered during the manufacture of wood pulp. The organic material is carbonized and then activated under controlled conditions to obtain active carbon grades of the desired form and adsorptive capacity. Aqua Nuchar A is recommended for the removal of tastes and odors from potable water supplies, to control odors at sewage plants and to correct industrial pollution. It has a phenol value of 20 mg/l (plus or minus 10 percent) or less as measured by the phenol test of the American Water Works Association. Available only in powdered form, a typical wet-sieve analysis gives the following minimum percentages passing each sieve: 100 mesh, 99 percent; 200 mesh, 98 percent; and 325 mesh, 95 percent. It has a surface area of 600 m^2/g and an apparent bulk density of 13.5 - 15 lb/ft^3. Ash content is 6 percent and moisture content is 3 percent.

Coagulation Studies

For successful operation of the high-rate clarification units comprising the pilot plant, it was essential that an efficient and effective coagulation system be developed. The floc particles had to settle rapidly, be strong enough to withstand the repeated cycling through the flocculation paddles, entrap a high percentage of the carbon and raw sewage particulate matter, and dewater readily. In addition, it was suggested inorganic chemicals should be avoided, if possible, to minimize the accumulation of inorganic matter in the carbon. Thermal regeneration and reuse of the carbon were not included in this study, but the process should minimize carbon contamination with inert solids.

Organic polyelectrolytes were tested as primary flocculating agents, augmented as indicated with various amounts of inorganic materials. A few experiments established the dosages of inorganic flocculants to support the balance of experiments.

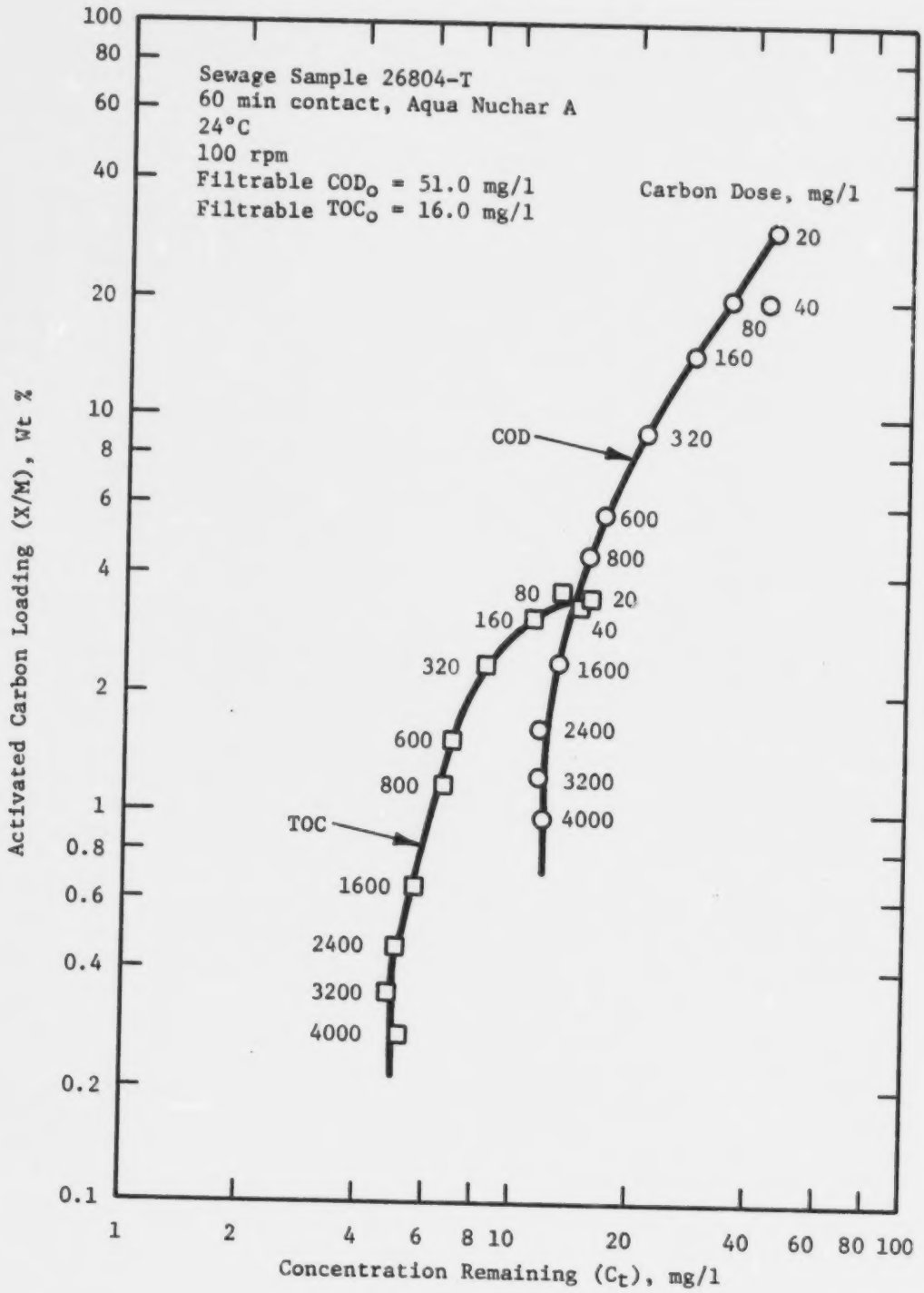

Figure 3: Extended Range Isotherm

The emphasis for much of the coagulation investigation was directed at cationic polyelectrolytes. A wide assortment of materials was available initially and promising ones were retested with factory-fresh supplies to eliminate effects of limited shelf life.

Products utilized and their respective manufacturers are as follows:

1.	Aquafloc	412	Dearborn Chemical Division
2.	CAT-FLOC	-	Calgon Corporation
3.	Magnifloc	521-C	American Cyanamid Company
4.	Magnifloc	560-C	American Cyanamid Company
5.	Primafloc	C-3	Rohm & Haas
6.	Primafloc	C-5	Rohm & Haas
7.	Primafloc	C-6	Rohm & Haas
8.	Primafloc	C-7	Rohm & Haas
9.	Purifloc	A-21	The Dow Chemical Company
10.	Purifloc	A-22	The Dow Chemical Company
11.	Purifloc	C-31	The Dow Chemical Company
12.	Purifloc	C-32	The Dow Chemical Company
13.	Purifloc	N-12	The Dow Chemical Company
14.	Triton	CF-32	Rohm & Haas
15.	Triton	QR-419	Rohm & Haas

Filter alum, ferric sulfate, ferrous sulfate, ferric chloride, activated silica, sodium hydroxide and lime were the inorganic materials utilized.

Over 800 individual jar tests were conducted on unfiltered screened and degritted raw sewage grab samples with variations in mixing speed, mixing time, carbon dosage, coagulating agent(s), order of coagulating agent and carbon addition, and the use of floc from previous tests.

Aliquots, generally 200 ml of sample, were treated first with activated carbon stock solution to either 200-mg/l or 250-mg/l dosages. While being stirred at test speed, the coagulating agent(s) were added, generally in the range of 2 to 20 mg/l. Stirring time was normally 20 minutes; however, this was shortened to 5 minutes if floc breakup was observed during the longer period. Stirring speed was 50 to 70 rpm with no flash mixing practiced. Supernatant turbidities were determined on most jars showing promise. A 5-minute settling period was allowed before sampling the supernatant via pipetting from beneath the liquid surface for turbidity measurement. Notes were made on the time required for the prominent floc particles to settle the 3.5-in. liquid depth to the jar bottom.

A substantial portion of these studies were undertaken before a carbon brand selection was finalized, so both Aqua Nuchar A and Darco S-51 were tested. Better floc formation, settling rate and completeness of clarification, perhaps because of higher bulk density (27-33 lb/ft^3)

and larger particle size (70% passing a 325-mesh sieve) were reported for the Darco S-51.

Each clarifier of the pilot plant was expected to exhibit individual coagulant requirements. A jar-test procedure simulating the two stages of treatment was utilized in about 20 percent of the aforementioned studies. Briefly, the procedure involved paired sets of jars. One of each pair representing the 1st stage or upstream clarifier and the other the 2nd stage or downstream clarifier. Treatment was tested in consecutive cycles. Each cycle for the 1st stage involved decanting the clarified supernatant, discarding a portion of the existing carbon slurry sediment, replacing this with a similar portion from the second-stage jar, adding raw sewage, mixing and coagulating. Each cycle of the 2nd stage involved decanting the previously clarified supernatant (sometimes for test, otherwise to waste) removing a portion of the previously formed carbon sediment (to 1st-stage jar), adding the 1st-stage decanted supernatant, mixing, adding the carbon and coagulating. In this fashion the requirements for flocculating the raw sewage suspended solids in the presence of already flocculated carbon were confined to the 1st-stage jar. Requirements for flocculating new carbon were confined to the 2nd-stage jar. Both jars retained and reused in subsequent cycles a quantity of previously-formed carbon-bearing floc for as many as nine cycles, during which floc characteristics change and coagulant dosage changes may be evaluated. The solids-recycle procedure is particularly applicable to the type of solids contact units used in the pilot plant.

Typical data reporting for jar studies of this type indicates the amounts and types of materials used, as well as notes on floc formation, quiescent settling ability, and clarification results as JTU or, as being poor, fair or good. The resulting slurry volume existing at the end of the settling period, as well as the number of previously formed slurry volumes returned to each jar are also recorded.

None of the polyelectrolytes when used individually were consistently effective in reducing supernatant turbidity to the desired range of 10 JTU except in dosages of 15 mg/l per stage or greater. Four products, C-31, C-32, C-7 and CAT-FLOC were found to be superior to others tested. They were equally effective at reduced feed levels when 1st-stage coagulation was aided by 10 mg/l filter alum. The polymers appeared to be somewhat selective in coagulating activated carbon well while rejecting fractions of the raw sewage suspended solids which contributed the bulk of residual turbidities in the jar supernatants. Modest amounts of alum used in the 1st-stage jars improved supernatant turbidities substantially. Calculations show filter alum produces less precipitate weight per mg/l fed than the other commonly used inorganic coagulants. This possibly could be an important factor when regenerating spent carbon.

Based upon price and ease of handling, C-31 was selected as the primary flocculating agent. Later, studies with the pilot plant would incorporate auxiliary alum feeds if found desirable.

Having completed selection of Aqua Nuchar A activated carbon and C-31 polyelectrolytic coagulant, operation of the pilot plant was initiated.

PILOT PLANT PROGRAM

Apparatus and Procedures

An existing constant-rate pilot plant was modified and applied to this study. The pilot plant was constructed in 1968 at the INFILCO Division Test Facility adjacent to the Municipal Water Pollution Control Plant serving the City of Tucson, Arizona.

The plant included two series-connected JBAS® ACCELATOR clarifiers, a type of recirculating-slurry solids-contact unit which is the principal component of a pre-engineered water treatment system used extensively in the beverage industry. As depicted in Figure 4, the pilot plant schematic flowsheet, the JBAS ACCELATOR clarifier contains an inlet-mixing chamber, flocculators and a slurry-recycle impeller on a common shaft, a slurry concentrator, and a clarified water zone with effluent take-off system.

Certain features of the clarifiers were modified at the time of installation for a previous study. A submerged-orifice peripheral launder designed for 30,000 gpd was substituted for the 6,000-gpd original effluent-collection system. The concentrator rim was modified by adding a spout to extract solids directly from the recirculating carbon slurry. This made solids collection independent of the slurry pool level. These alterations were retained for this study. New modifications included reducing the concentrator volume from 17.4 gal. to 6.9 gal. by means of a liner and the addition of a peripheral aluminum scum baffle to the upstream unit.

Operation at throughput rates as high as 28,000 gpd was realized, resulting in an overflow rate of 1.5 gpm/ft^2 of clarification area based on the 12.9-ft^2 cross-section of the annular clarification zone.

Each of the 730-gal. clarifiers (4.5 ft diameter by 7.5 ft sidesheet) included a recirculation zone volume of 225 gal. which provided carbon contact times of 35, 23 and 12 minutes at proposed operating rates of 0.5, 1.0 and 1.5 gpm/ft^2, respectively.

In a solids-contact unit of this type, the throughput rate cannot exceed the pumping capacity of the recirculation system if short-circuiting to the solids separation zone is to be avoided. The estimated pumping capacity of the slurry recirculation impeller in the pilot units was five times the maximum operating rate. The recirculation flow also prevents deposition of solids on the tank floor. Except for material isolated in the slurry concentrator, all of the carbon inventory was maintained in suspension during operation.

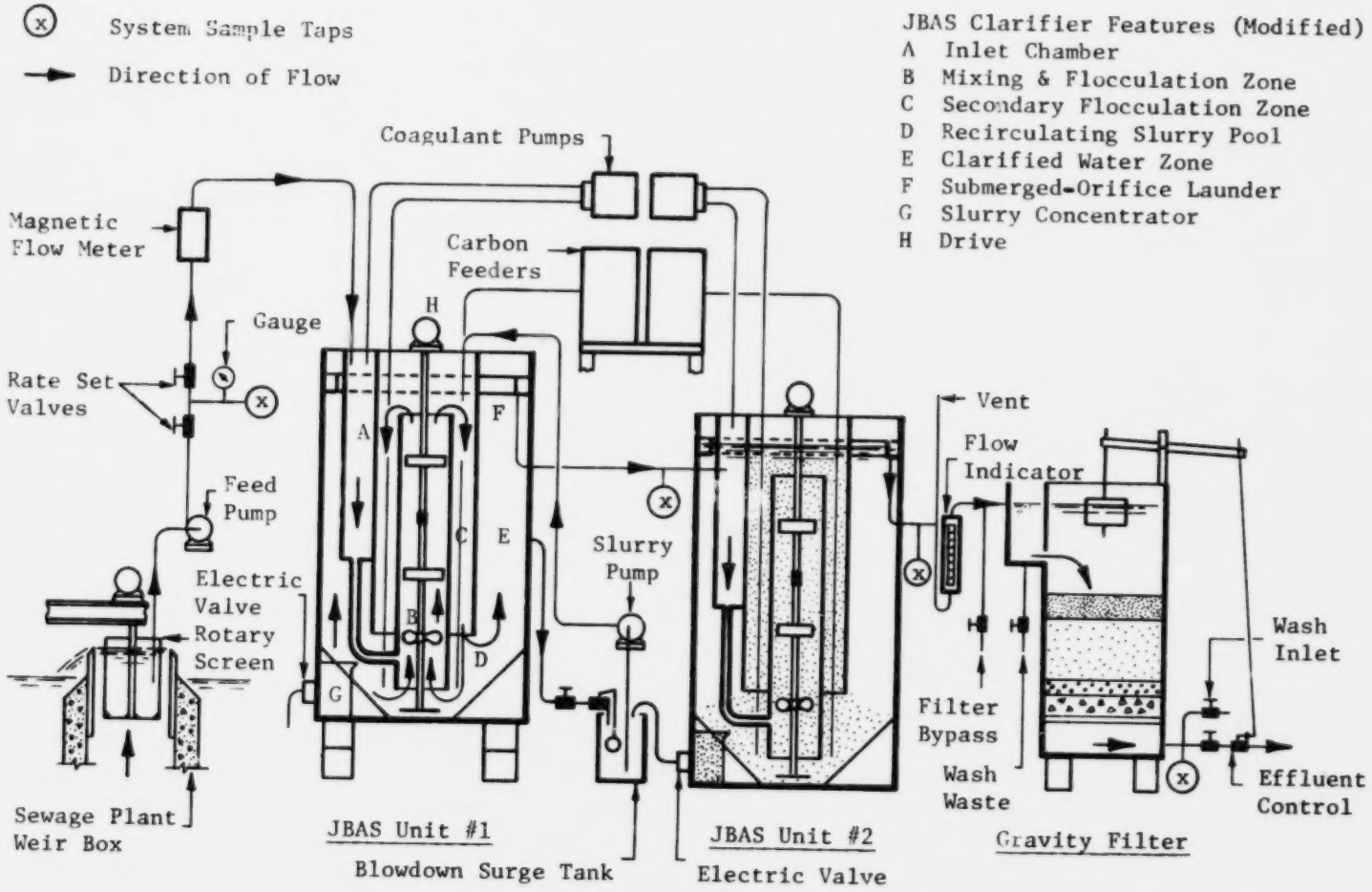

FIGURE 4: Schematic, Pilot Plant System

The pilot plant (Figure 5) was erected out-of-doors in an orientation which differed only slightly from that shown in Figure 4. The feed stream was pumped from the interior of a 1/4"-mesh screened revolving basket at the sewage plant weir box some 200 ft distant. This basket rejected raw sewage solids of a size and type that otherwise would foul the centrifugal feed pump, influent rate control valves and slurry discharge valves. Center-driven at 7 rpm for self-cleaning, the basket had 2.5 ft^2 of submerged screen area and a 6-in. freeboard at normal liquid level.

Two manually operated weir-type diaphragm valves in the feed line were throttled to set the throughput rate. The upstream valve was throttled to reduce the line pressure at the downstream valve which was adjusted to give the desired rate as indicated by a magnetic flowmeter. Two valves in series provided greater valve openings for sewage solids than a single one. The resilient diaphragm construction of these valves also provided a measure of self-clearing. Solids would pass that otherwise would clog a conventional valve. Occasionally one or both valves would require manipulation to dislodge solids.

The feed stream entered the inlet chamber of the upstream clarifier (Unit #1) discharging below liquid level to avoid foaming. Unit #1 effluent flowed by gravity to the inlet chamber of the downstream clarifier (Unit #2) and thence by gravity to the gravity anthracite-sand filter for final treatment.

Carbon feed was countercurrent to the throughput flow. Fresh powdered activated carbon was gravity fed by a volumetric slurry feeder to the bottom end of the inner draft tube of Unit #2 along with a separate feed of polymer flocculant. A second carbon feeder was used during initial start-up to permit simultaneous development of slurry in both units and as a standby for the other feeder.

Flocculated carbon slurry recirculating within Unit #2 was permitted to accumulate to a manageable concentration which was controlled by timer-actuated blowdown from the internal concentrator. The repeating-cycle timer used afforded independent adjustment of both the valve-open and valve-closed portions of the blowdown cycle from a fraction of a minute each up to a 2-hr total cycle time.

Used-carbon blowdown from this unit was collected in the 10-gal. polyethylene blowdown surge tank and, together with makeup flow of semi-treated wastewater from mid-depth in Unit #1, transferred continuously to the bottom of the inner draft tube of Unit #1. A stainless steel Jabsco Pump operating at low speed was used for the slurry transfer after a centrifugal pump and hydraulic ejector system was found to break up the floc. Polymer was also fed separately to this region.

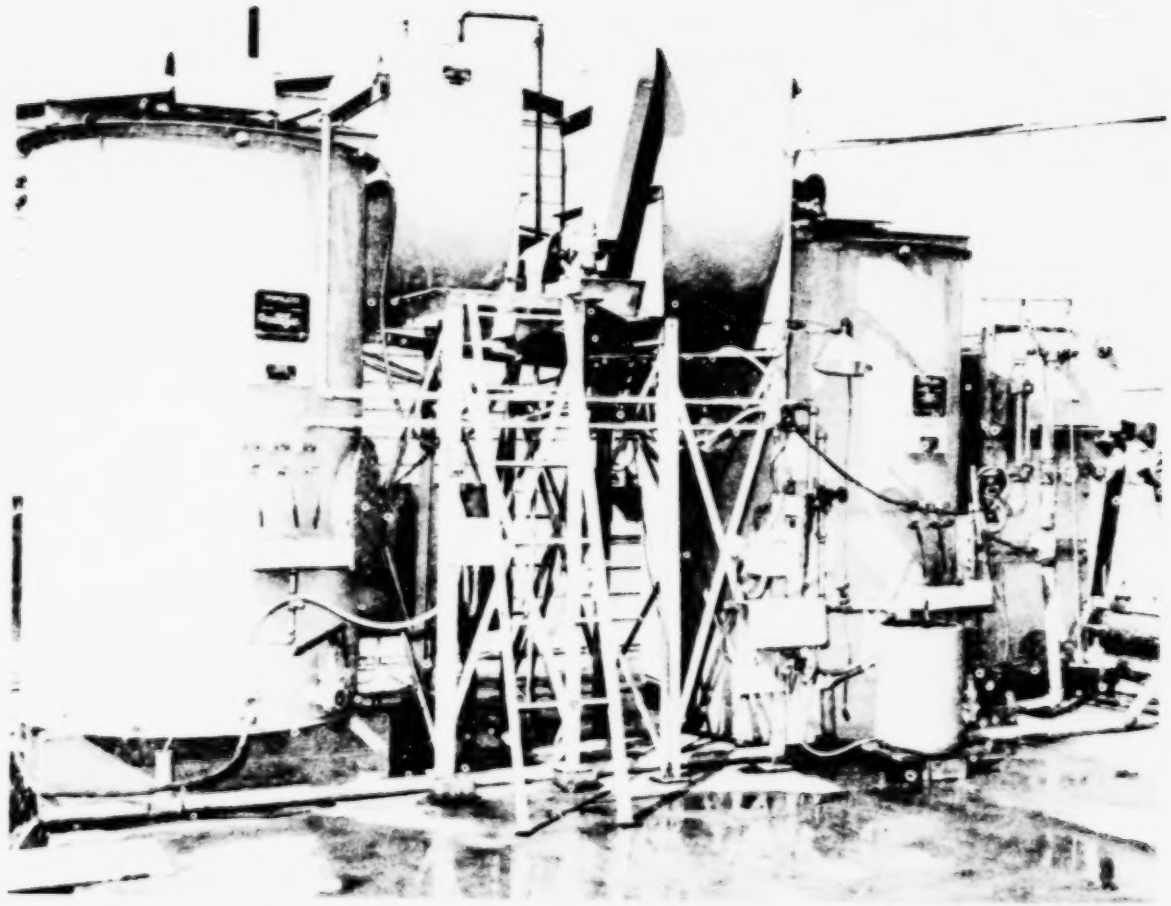

FIGURE 5: Photograph of 28,000 gpd Pilot Plant

Blowdown from Unit #1, also timer controlled, was to waste when not being measured or sampled.

Unit #2 effluent could be divided so that any portion or all of it could be wasted or directed to a 2-ft square by 6-ft high gravity filter. The filter was constructed of Transite and Lucite with two opposing transparent sidewalls allowing visual observation of the contents. It included an extruded asbestos-cement underdrain system supporting 11 inches of graded gravel, 16 inches of filter sand and 6 inches of anthracite coal. The filter sand had a specified effective size between 0.45 and 0.55 mm and a maximum uniformity coefficient of 1.75. The anthracite coal had a specified effective size between 0.90 and 1.20 mm and a maximum uniformity coefficient of 1.7.

A float-operated effluent control valve was added to the filter to pace filtration with plant flows up to 12 gpm. During higher plant operating rates, effluent from the #2 Unit was wasted to limit the filtration rate to 12 gpm (3 gpm/ft^2 of filter area) or less. A transparent plastic hose, tapped into the underdrain and extended to the top of the filter, provided head loss indication during operation.

As a matter of convenience, locally available well water pumped from a 2000-gal. storage tank was used for filter backwashing. The wash water was batch-chlorinated in the storage tank with commercial bleach or calcium hypochlorite prior to backwashing. The filter was backwashed with chlorinated water every day for biological growth control for the first 97 days. Backwashing at maximum head loss of 54 inches was adopted for the remaining 186 days of operation. The wash rate was set manually while observing the extent of sand-coal expansion. Wash water consumption was computed from tank draw-down. Wash rate information is presented elsewhere in this report as surface-wash procedures varied during the program.

Sample taps were provided on each contactor-clarifier unit for slurry sampling and determination of the slurry-clear water interface elevation. A specially designed multiple-dipper sampler, supplied with continuous streams of Unit #1, Unit #2 and filter effluents automatically composited samples of these streams. The dippers collected aliquots at 1-minute intervals throughout the 24-hour compositing periods. Samples of the feed stream were collected manually each hour and sewage plant effluent samples were collected manually every two hours for these 24-hour composites. The samples were accumulated in polyethylene containers refrigerated to a temperature between 1° and 3° C.

The pilot plant program was conducted in a flexible fashion such that sequential operating period objectives could be adjusted as results became available. New data were reviewed frequently in consultation with the Project Officer. The program has been divided into seven consecutive phases:

1. Brief operation of the system at 6.5 gpm (0.5 gpm/ft^2) and a carbon dosage of 145 mg/l for equipment shakedown and operator training. This flow was selected as a practical minimum dictated by previous experience as well as difficulties associated with controlling the flow rate of the solids-bearing feed stream. Sampling and analytical work were phased in simultaneously. Operation continued through Operating Periods 1, 2, 3 and 4 during which time the flow rate was increased incrementally to the maximum of 19.4 gpm (1.5 gpm/ft^2) while holding all other controlled variables constant.

2. Operation at 19.4 gpm and 145 mg/l carbon feed during sequential reduction in polymer dosage to Unit #1 from 20 mg/l to 6 mg/l, reduction in polymer dosage to Unit #2 from 15 mg/l to 6 mg/l and initiation of 10-mg/l alum dosage to Unit #1. This phase consisted of Operating Periods 5A, 5B, 5C, 6A and 6B.

3. Brief operation duplicating Operating Period 6B except substituting Darco S-51 activated carbon for Aqua Nuchar A. The initial Operating Period 7 was halted to modify the plant to better manage the type of blowdown solids produced. Operating Period 7A was with the modified system.

4. Operation again duplicating Operating Period 6B with Aqua Nuchar A to bracket the S-51 run and monitor any significant datum shift. The modifications added for Period 7A were removed. This was Operating Period 8.

5. Sequential Operating Periods 9 through 11C during which the flow rate was reset to 13.0 gpm and polymer dosages to both units was varied from 4 mg/l per unit to 6 mg/l per unit as alum dosage to Unit #1 was varied from 0 to 25 mg/l and alum dosage to Unit #2 was studied at 0, 25 and 20 mg/l.

6. Operation at 13.0 gpm (1.0 gpm/ft^2) through Periods 12A, 12B and 12C attempted to avoid the use of polymer while feeding alum at dosages of 20 mg/l to 50 mg/l per unit. Polymer at dosages of 2 mg/l to 3 mg/l per unit was required.

7. Operation at 13.0 gpm with carbon dosages of 200 mg/l, 250 mg/l and 100 mg/l was studied during Operating Periods 13, 14 and 15. Polymer and alum were fed to both units in amounts found necessary for acceptable clarification.

During each operating period, composite and grab samples were collected for development of laboratory isotherms and analysis sufficient to define process performance, i.e., pH, color, turbidity, suspended solids,

filtrable and unfiltered COD, filtrable and unfiltered TOC, unfiltered BOD_5 and slurry blowdown solids concentrations. Other data recorded included quantities and types of treatment chemicals; plant flow; rate and volume of blowdown; slurry volume after 5 minutes of settling; slurry level; filter head loss; and the filter backwash frequency, rate and duration. Still other data were accumulated on a lesser scale and are discussed separately. Records were also made of events which interfered with or otherwise affected plant operation or performance.

Fourteen-day runs at optimum conditions for the situation under test were programmed to avoid biasing data because of weekly cycling of the feed characteristics. In fact, few of the periods went 14 days or longer as physical treatment criteria under study were frequently demonstrated in less time. Other periods were halted prematurely by mishap or the number of days of data originally collected during a run was reduced later by discarding after reviewing records of operation and events. Continuous plant operation was undertaken to parallel the reference sewage treatment plant situation and also to avoid potential biological activity interferences that might occur with intermittent operation.

Experimental Results

A sizeable quantity of raw analytical and operational information was accumulated during the pilot plant operation. Except when presented in example Figures and Tables, raw data are omitted from this report for the sake of brevity. Mean-data values were computed for individual operating periods and are the basis for experimental results reported. Probability analyses were not attempted as too few bits of information were retained from some periods to justify this approach.

Evaluation of the period-mean results is aided by graphical presentation of the measured variables vs. operating period from beginning to end of the field program.

Phase 1 - Initial operation of the pilot plant at 0.5 gpm/ft^2 clarifier overflow rate continued for 28 days. Operator training, equipment shakedown, initiation of sampling and analytical laboratory programming consumed the early portion of this period. Valid information was collected during the later portion. Table 2 presents the program followed and results obtained for this and the other operation periods comprising Phase 1. A high polymer feed was used to assure flocculation and 145 mg/l Aqua Nuchar A (ANA) activated carbon feed was the computed dosage required to achieve 50 percent filtrable COD removal by two-stage countercurrent adsorption with assumed values of C_o = 100 mg/l COD and $1/n = 1.52$. These assumptions originated from the many grab-sample isotherms developed earlier.

TABLE 2

Phase 1, Pilot Plant Performance by Operating Period, Mean Values

	Operating Period			
	1	2	3	4
CONTROLLED VARIABLES				
Overflow Rate, gpm/ft^2	0.5	0.75	1.0	1.50
Carbon Feed, mg/l ANA	145	145	145	145
C-31 Feed, mg/l Unit #1	20	20	20	20
Unit #2	20	15	15	15
Alum Feed, mg/l Unit #1	0	0	0	0
Unit #2	0	0	0	0
No. of Days Sampled	17	10	7	18
No. of Operating Days	28	13	7	27
PILOT PLANT REMOVALS				
TOC unfiltered, %	83.8	86.1	83.1	86.6
TOC filtrable, %	63.1	64.8	55.5	66.1
COD unfiltered, %	90.6	91.0	90.0	89.6
COD filtrable, %	65.8	69.6	61.0	69.5
BOD_5 unfiltered, %	92	89	87	90
Sus. Solids, %	98.6	96.7	96.2	97.8
SEWAGE PLANT REMOVALS				
TOC unfiltered, %	82.9	72.5	84.5	90.1
COD unfiltered, %	87.3	78.0	88.8	91.0
BOD_5 unfiltered, %	81	66	92	88
Sus. Solids, %	95.1	84.0	91.6	95.2

Filtrable COD of the feed stream exceeded 100 mg/1 (Figure 6) during the entire Phase; however, removals between 61.0 and 69.6 percent were recorded. Pilot plant effluent COD was less than the 50-mg/1 level suggested as the original target value. Filtrable TOC removals were slightly lower, ranging from 55.5 to 66.1 percent. Filtrable COD carbon loading (X/M) reached 54 percent during Periods 2 and 4.

Blowdown from Unit #1 was excessive at start-up amounting to 13 percent of the feed stream. Hydraulic currents were scouring the slurry concentrators so the concentrator spouts were eliminated by filling them with grout. Slurry would now feed to the concentrators by decanting from the upper reaches of the slurry pool and by outer draft tube discharge impinging upon the filled spout and flowing horizontally over it. Blowdown from Unit #1 decreased to 2.5 percent after restarting the plant. Unit #2 blowdown decreased from 3.2 to 1.0 percent of throughput as a result of the alteration.

From the onset of operations it was obvious the raw sewage solids captured in Unit #1 slurry would require discharging 2 to 3 times more blowdown from this unit than from Unit #2. As the operating rate was increased, dewatering of the slurry in the concentrator diminished because of the shortened retention times. This situation was aggravated by slurry deflocculation originating in the centrifugal pump and ejector carbon-advance system such that, in Operating Period 4 at the 1.5-gpm/ft^2 overflow rate, slurry could not be conditioned well enough for control. The slurry interface rose well above the tap in Unit #1 that supplied water to the centrifugal pump suction and as make-up water to the carbon advance surge tank. In effect, the system was recycling up to 7.5 gpm of damaged floc through Unit #1. This system was replaced with the resilient-impeller pump described earlier, for the data collecting portion of Period 4.

The diurnal variation in feed suspended solids caused the slurry inventory to cycle widely each day. Draft tube slurry concentration (Vol %/5 min settling time) would build steadily through the afternoon and evening hours and decrease quite rapidly in the morning hours. The program to maintain approximately 10 to 15 Vol %/5 min throughout the day required frequent adjustment of the blowdown-system timers. As experience was accumulated, the program shifted to permitting the normal cycle to occur with timer adjustments made only if the tests indicated significant deviation from the expected value for that time of day. Generally the cycle range for Unit #1 was between 5 and 15 Vol %/5 min. The cycle range in both units could be varied, however, as the proportions of, and types of, coagulants were varied. During Phase 1, Unit #2 draft tube solids varied only 1 to 2 Vol %/5 min per day which approaches the reproducibility limits of the test. There was a tendency to ignore a gradually increasing or decreasing trend overly long and then overcorrect timer adjustment. Figure 7 is representative of draft tube slurry data field records for this period.

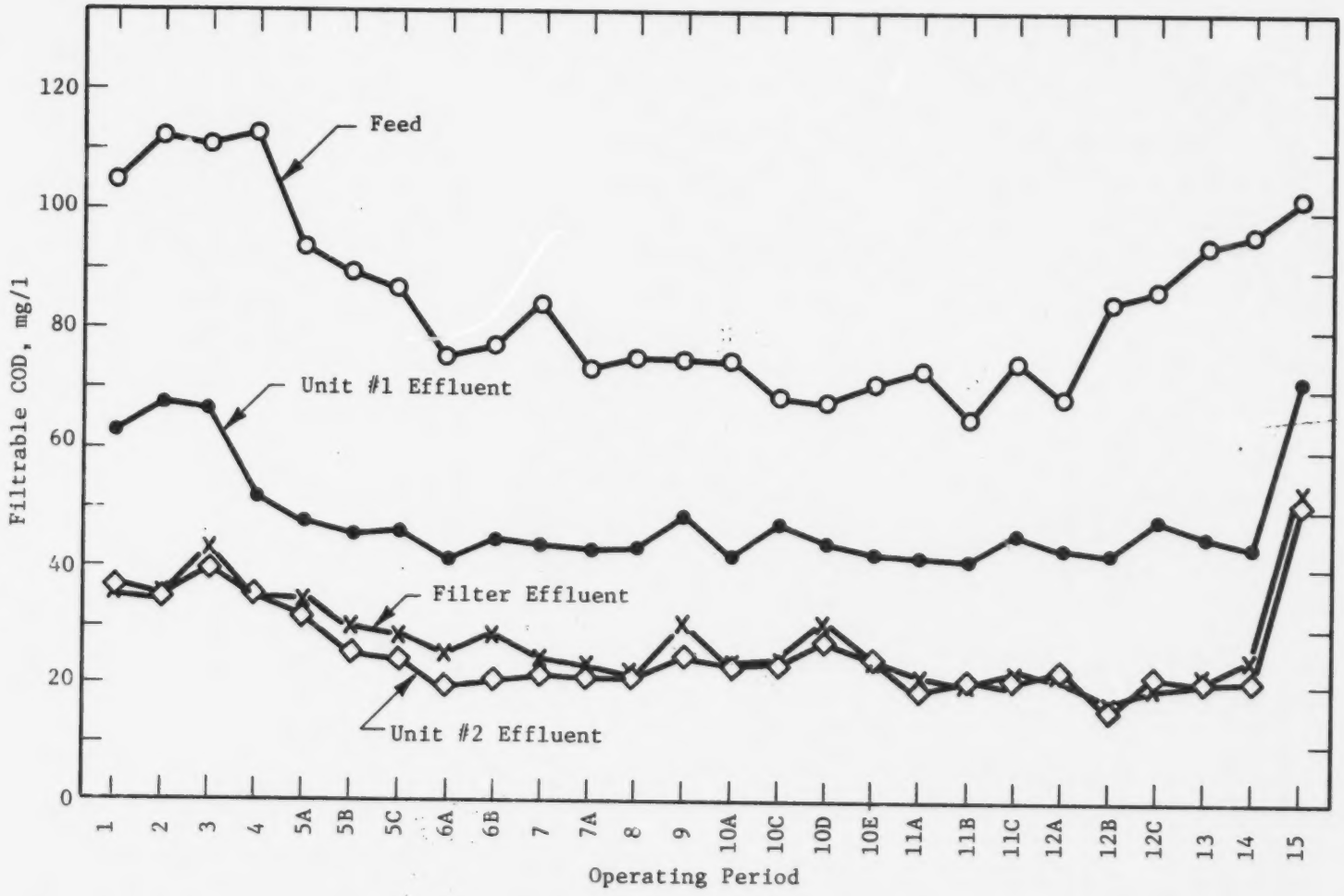

FIGURE 6: Mean Values of Filtrable COD

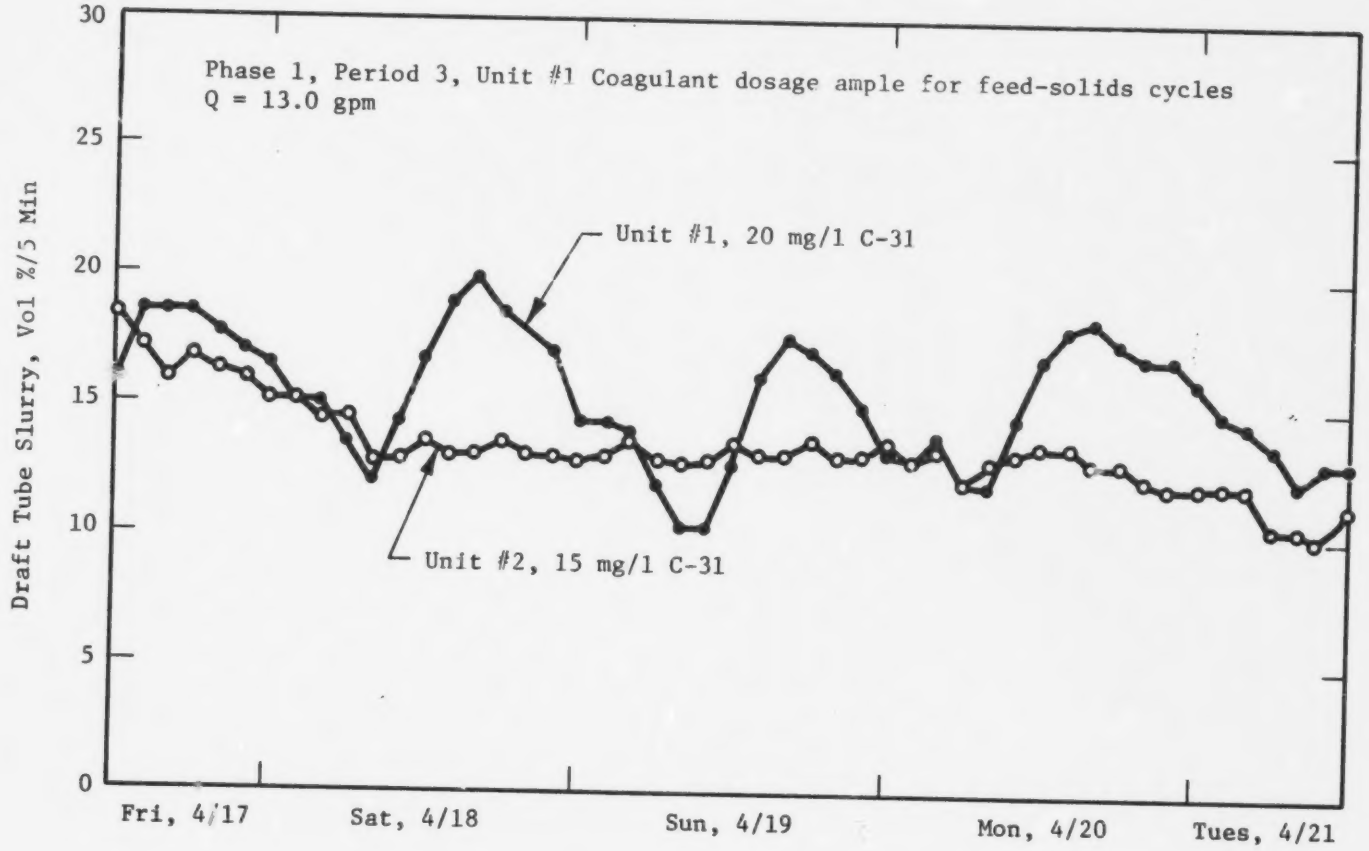

FIGURE 7: Diurnal Variation, Clarifier Slurry Settling Tests, Phase 1

Having established pilot plant operation to the hydraulic capacity limits of the design while simultaneously delivering an effluent comparable to the parallel sewage treatment plant this portion of the program was terminated.

Phase 2 - At the quoted price of 31¢/lb for C-31 in tankcar quantities, Phase 1 treatment cost 9¢/1000 gal. of sewage treated for polymer alone. Performance of the pilot plant at reduced polymer dosages, augmented with a minimum effective alum dosage, was investigated over the next 35 days during Operating Periods 5A, 5B, 5C, 6A and 6B. Overflow rate was held at 1.5 gpm/ft^2 and activated carbon feed was maintained at 145 mg/l ANA.

Physical performance of the clarifiers had been marginal. Adversely affected by factors already reported, plus several power outages, storm flooding of the inlet screen and other incidents normal to plant shakedown, the pilot plant had demonstrated that relatively high suspended solids carryover could be tolerated. Clarifier effluent quality, in terms of carryover, could be sacrificed if cost reduction could be demonstrated. Arbitrary acceptable carryover levels of 50 mg/l suspended solids for Unit #1 and 25 mg/l suspended solids for Unit #2 were adopted as guide lines. Figure 8 shows the suspended solids relationships for the various sample streams for the program.

The stepwise reduction in polymer to 6 mg/l per unit while feeding 10 mg/l alum to Unit #1 is listed in Table 3 along with mean performance data for these five periods. Figure 8 shows solids carryover from the clarifiers was within limits; however, feed solids also decreased during this time (May and June) such that coagulant requirements would be naturally lower. Filtrable TOC and COD of the feed dropped significantly during the same period (Figures 6 and 9); however, removal percentages tended to improve slightly (Figures 10 and 11). These Figures also show the dramatic shift in removal capability that occurred in the previous phase as the result of the improved carbon advance system. Unit #1 was now responsible for approximately 75 percent of the adsorbate removal as compared to 55 percent previously. As the feed stream TOC and COD diminished during Phase 2, lowering the adsorption potential in the upstream clarifier, the adsorbate removal division equalized substantially.

Five-day BOD removals in the tables represent order-of-magnitude values only. Insufficient determinations were made to establish accurate mean information. Five-day BOD determinations (Figure 12) were usually done twice weekly.

Clarifier blowdown remained reasonably constant at 6.5 to 7.0 percent for Unit #1 and 3.0 percent for Unit #2 throughout the period (Figure 13). Blowdown samples contained twice the suspended solids measured in the recirculating slurry. Unit #1 slurry averaged 2.63 g/l vs. blowdown

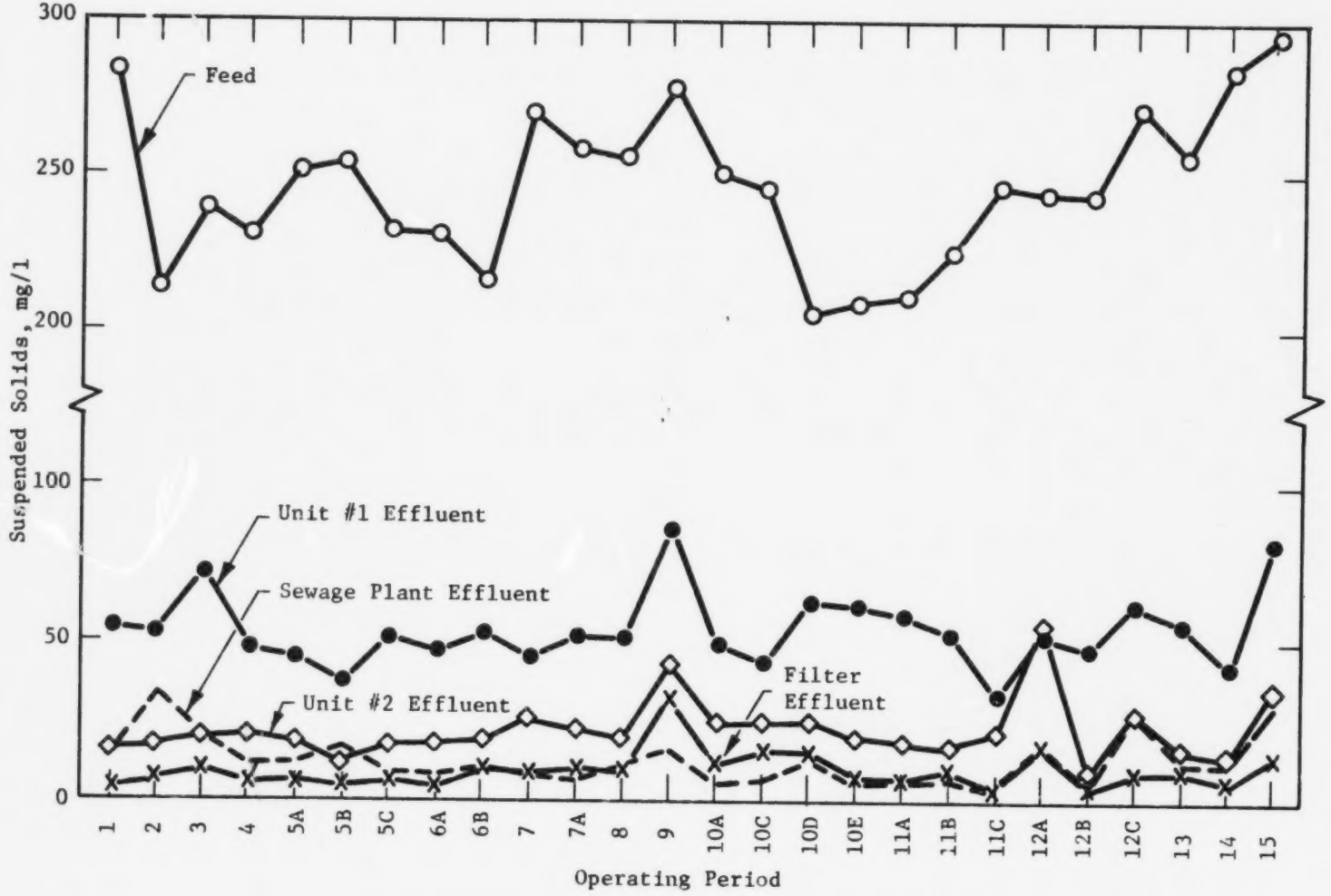

FIGURE 8: Mean Values of Suspended Solids

TABLE 3

Phase 2, Pilot Plant Performance by Operating Period, Mean Values

	Operating Period				
	5A	5B	5C	6A	6B
CONTROLLED VARIABLES					
Overflow Rate, gpm/ft^2	1.5	1.5	1.5	1.5	1.5
Carbon Feed, mg/l ANA	145	145	145	145	145
C-31 Feed, mg/l Unit #1	20	15	12.5	12.5	6
Unit #2	15	15	12.5	6	6
Alum Feed, mg/l Unit #1	10	10	10	10	10
Unit #2	0	0	0	0	0
No. of Days Sampled	5	4	16	4	6
No. of Operating Days	5	4	16	4	6
PILOT PLANT REMOVALS					
TOC unfiltered, %	82.2	86.5	88.5	91.8	88.5
TOC filtrable, %	60.5	66.8	61.8	62.3	64.0
COD unfiltered, %	89.6	90.3	88.7	89.7	85.2
COD filtrable, %	62.7	67.1	66.6	67.2	62.8
BOD_5 unfiltered, %	91	94	89	-	91
Sus. Solids, %	97.6	98.3	97.4	98.3	95.4
SEWAGE PLANT REMOVALS					
TOC unfiltered, %	84.8	84.1	89.0	90.7	88.8
COD unfiltered, %	90.0	86.1	90.0	89.6	88.5
BOD_5 unfiltered, %	91	94	89	-	90
Sus. Solids, %	96.6	93.2	96.6	96.5	95.4

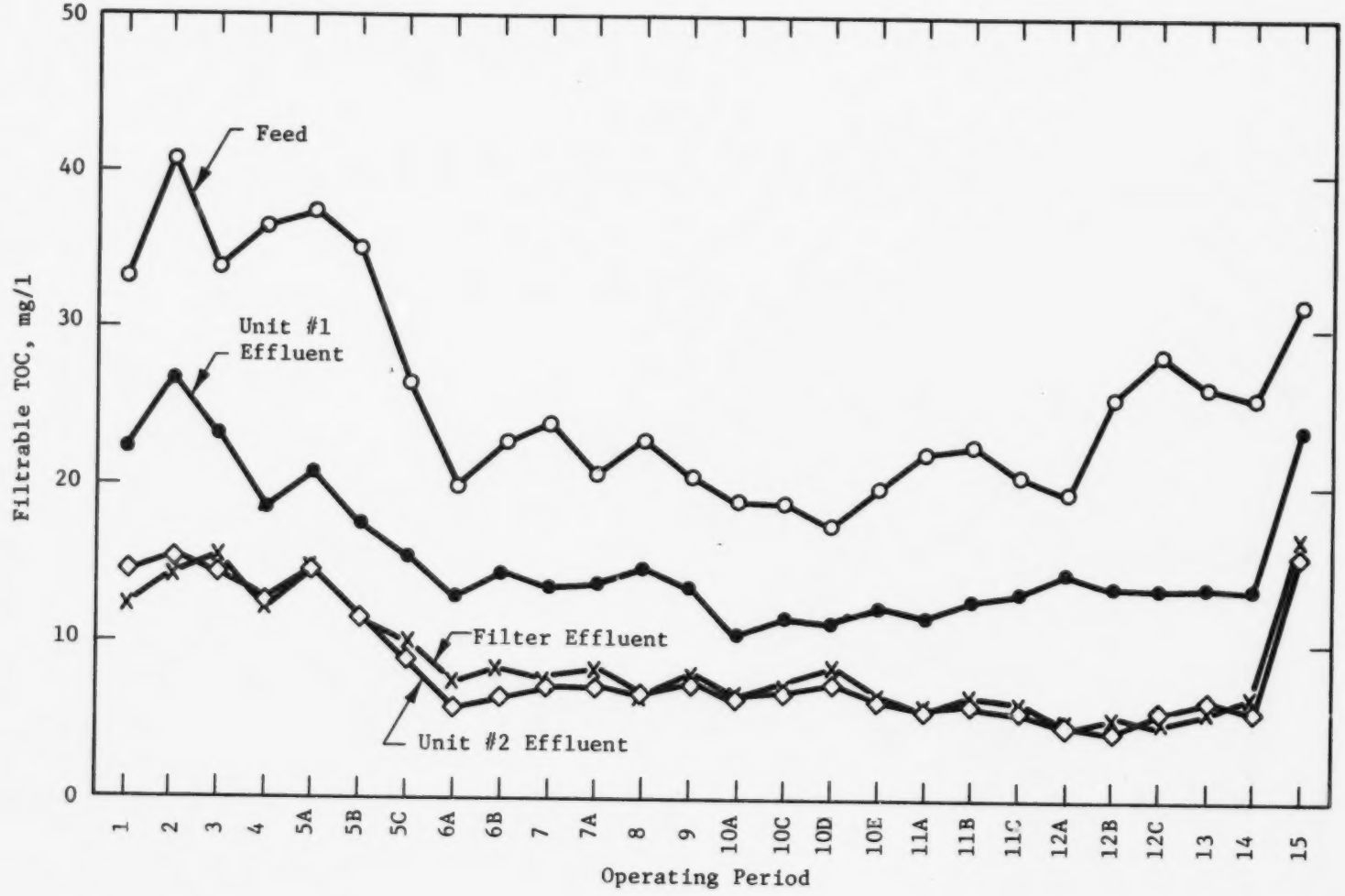

FIGURE 9: Mean Values of Filtrable TOC

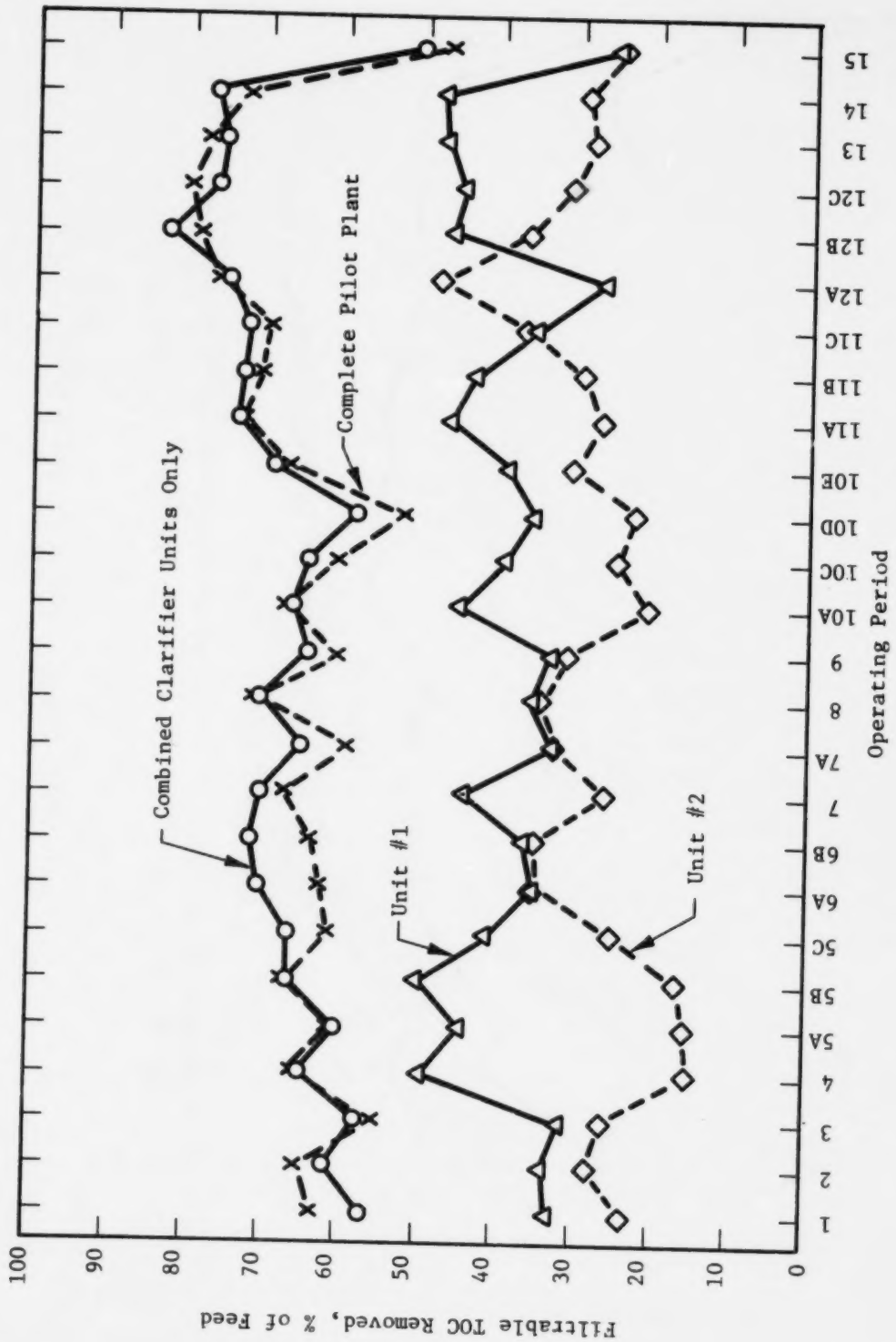

FIGURE 10: Mean Values of Filtrable TOC Removal

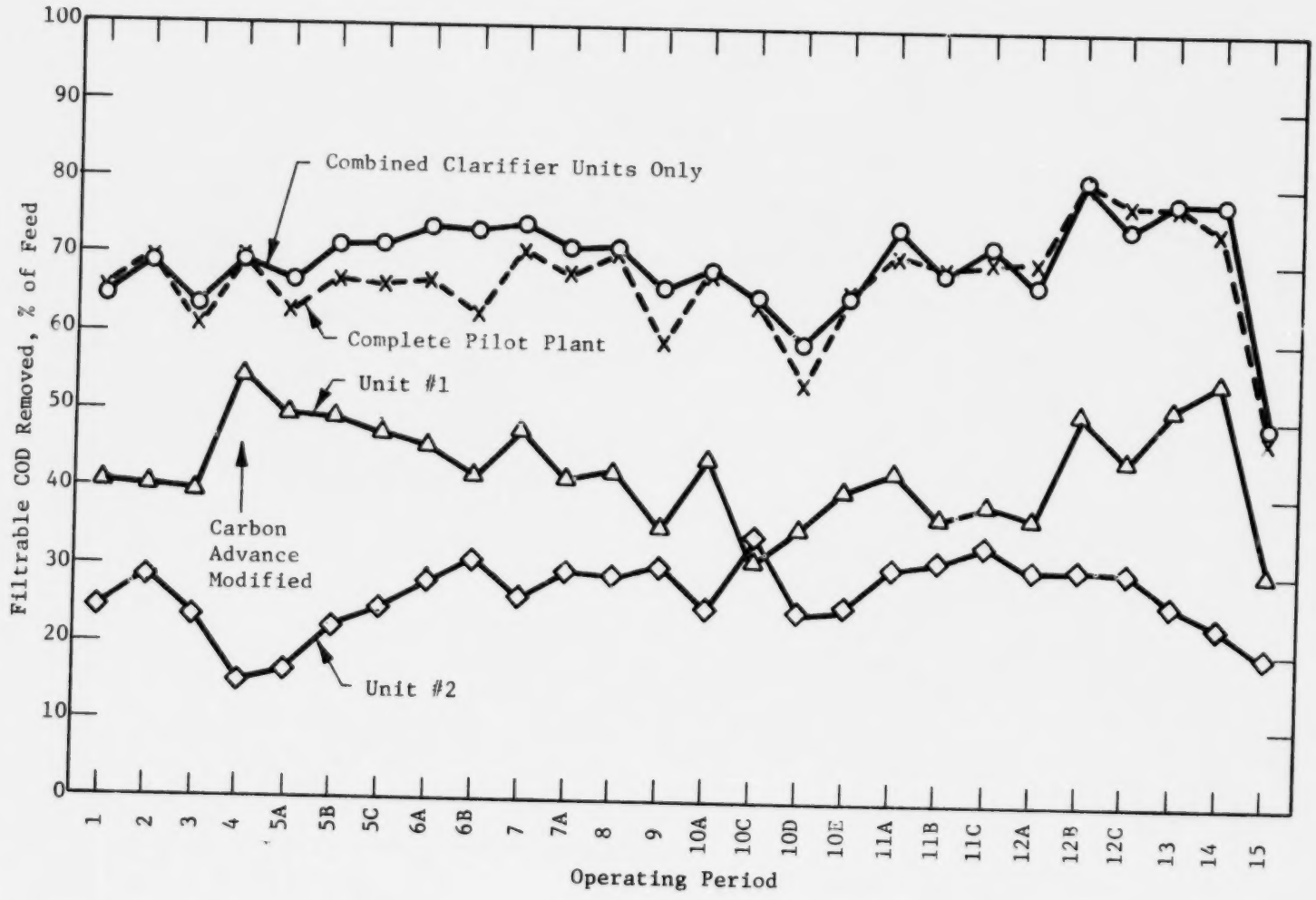

FIGURE 11: Mean Values of Filtrable COD Removal

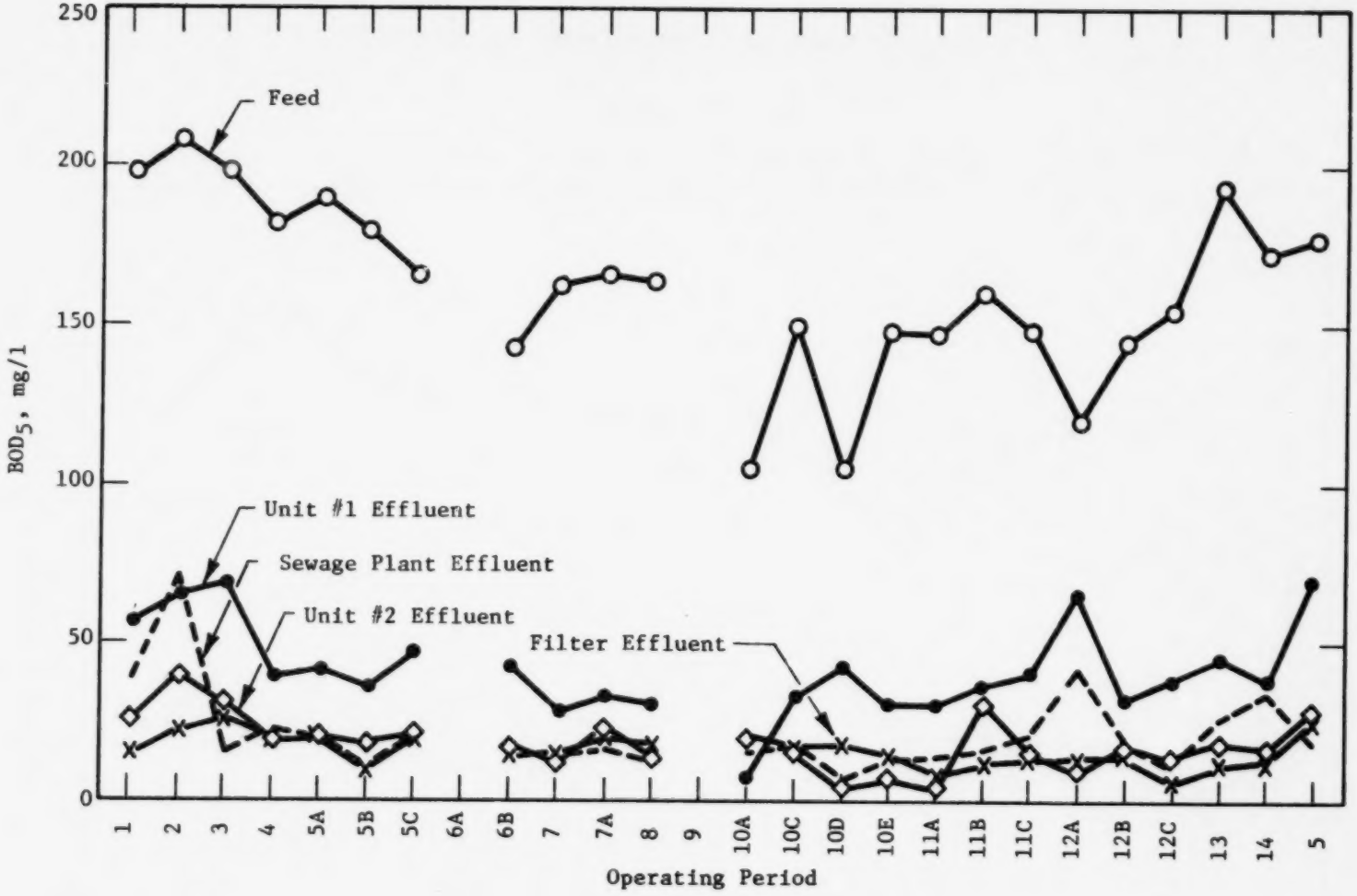

FIGURE 12: Mean Values of BOD_5

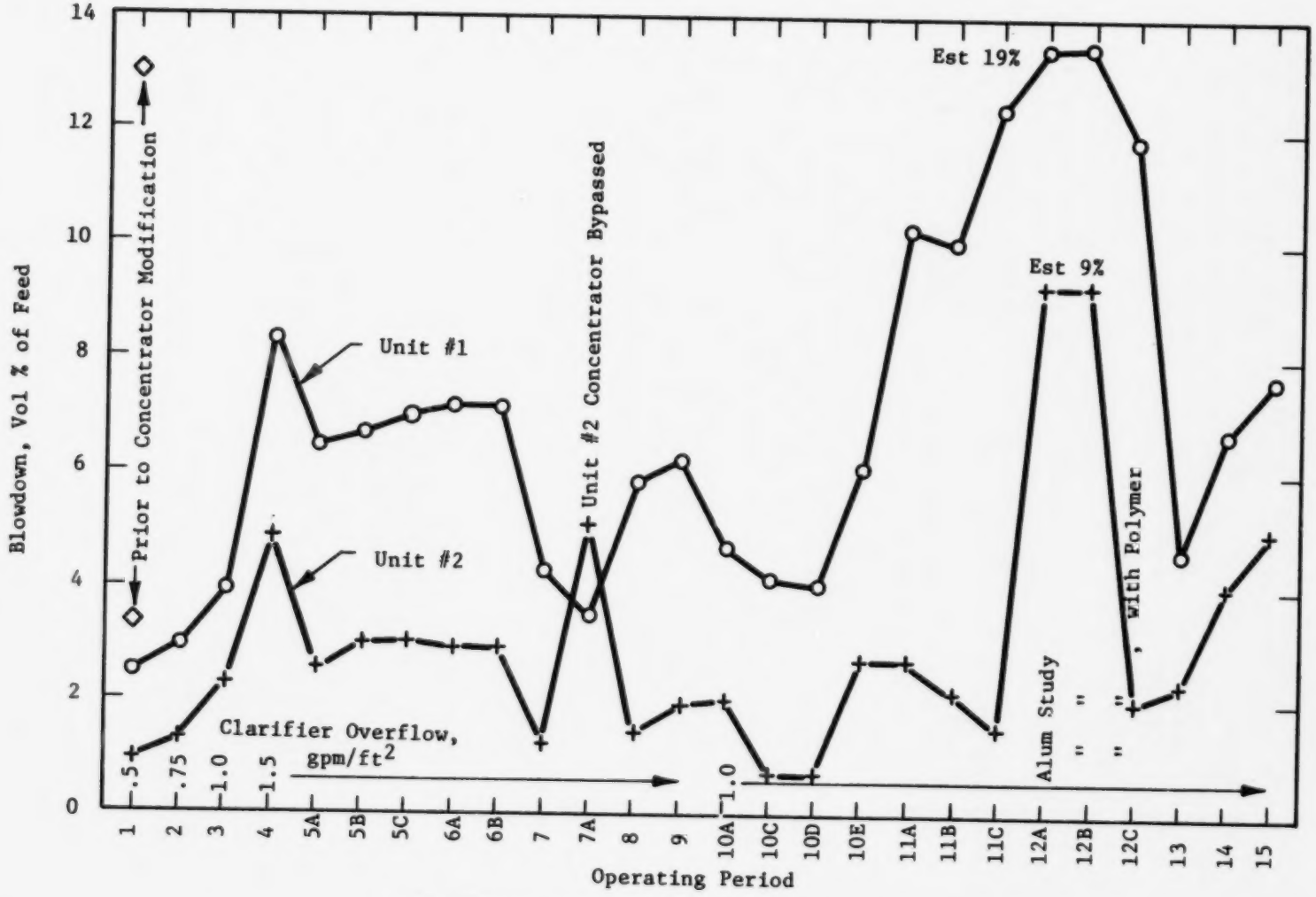

FIGURE 13: Mean Values of Clarifier Blowdown

of 6.15 g/l. Unit #2 slurry averaged 4.29 g/l vs. blowdown of 8.52 g/l. Independent dewatering tests (Figures 14 and 15) had already confirmed slurry solids content could be increased by a factor of 10 with 10 minutes settling time. The pilot plant concentrators were undersized for the quality and quantity of floc produced but physical alterations were deferred in anticipation of future operations at reduced flow rates.

During Operating Period 5C gravity filter operation was revised. Heretofore, filter runs had been purposely terminated at 24 hours for backwashing with growth-inhibiting chlorinated wash water. Modest amounts of algae and slime could be seen on the supporting gravel visible through the Lucite side walls. Covering these areas with opaque aluminum sheet had limited algae growth to a static condition. Filter effluent turbidities had persisted at 10 to 13 JTU (Figure 16). Examination, as in previous studies[6], indicated the material was not powdered carbon but the products of biological growth. The mat left on filter discs was green-brown colored.

Filter head loss at a surface loading of 2.6 gpm/ft^2 increased from a clean filter reading of 5 in. to 33 in. at the end of 24 hours. Volume filtered per day was 15,000 gal. with the balance of clarified liquid lost through blowdown and the filter bypass system.

After the procedure was revised to extend filter runs to the maximum head loss of 54 in. (operator duties permitting), runs averaged 28, 31 and 30 hours for Operating Periods 5C, 6A and 6B, respectively. Head loss was recorded hourly throughout the program (Figure 17). Mean filter effluent turbidity increased to 19 JTU for Period 6B. Nighttime backwashing (during which it was difficult to distinguish the coal-wastewater interface), extended periods between chlorination and higher bed differentials all are assumed to contribute to the solids discharge.

The mean washwater consumption was 647 gal. per wash or 4.3 percent of the volume filtered at this filter rate. In backwashing, the operators were instructed to brush down the filter freeboard as the unit was decanting to wash level. A low upflow rate, sufficient to barely expand the 6-in. coal bed, was established and the coal raked vigorously to dislodge the accumulated carbon-polymer floc. This was followed by a 5-minute or longer backwash at 36 percent (8 in.) bed expansion. The stored wash water was warm, up to 90°F at times, and instantaneous wash rates of 20 gpm/ft^2 were recorded.

Raking the filter media simulated surface washing, a step deemed essential to eliminate carbon-polymer-filter media mud balls noted in earlier studies. No mud balls developed.

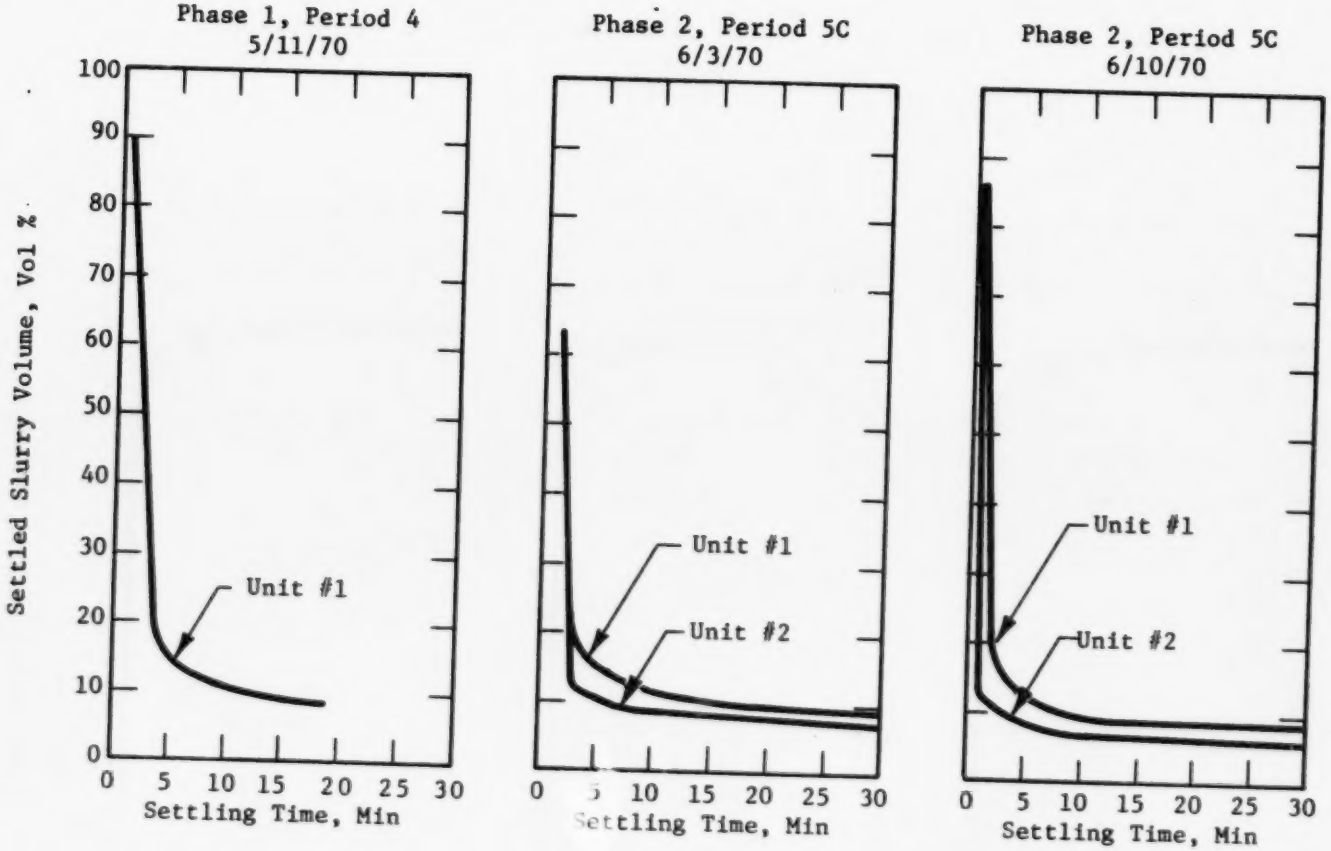

FIGURE 14: Recirculating Carbon Slurry Settling Rate Curves, Part 1

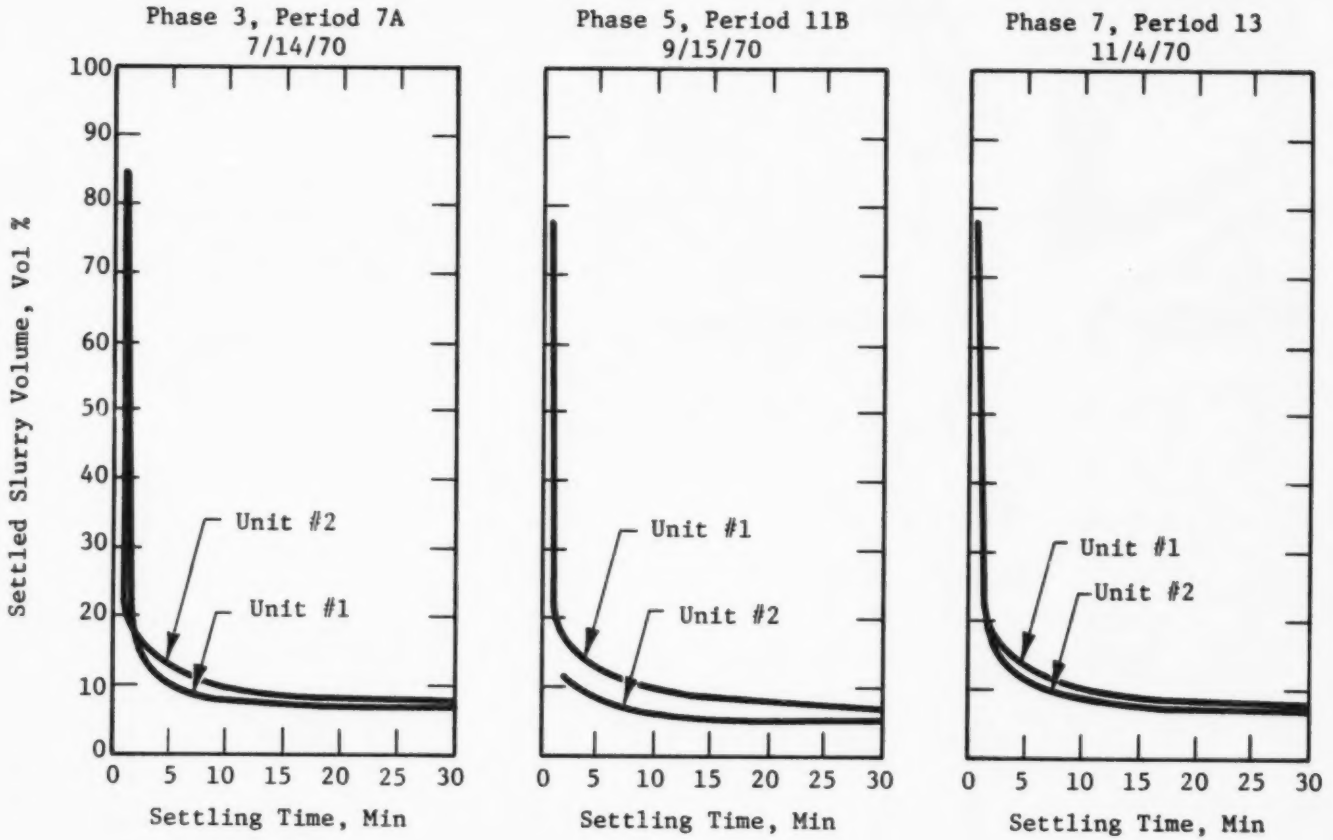

FIGURE 15: Recirculating Carbon Slurry Settling Rate Curves, Part 2

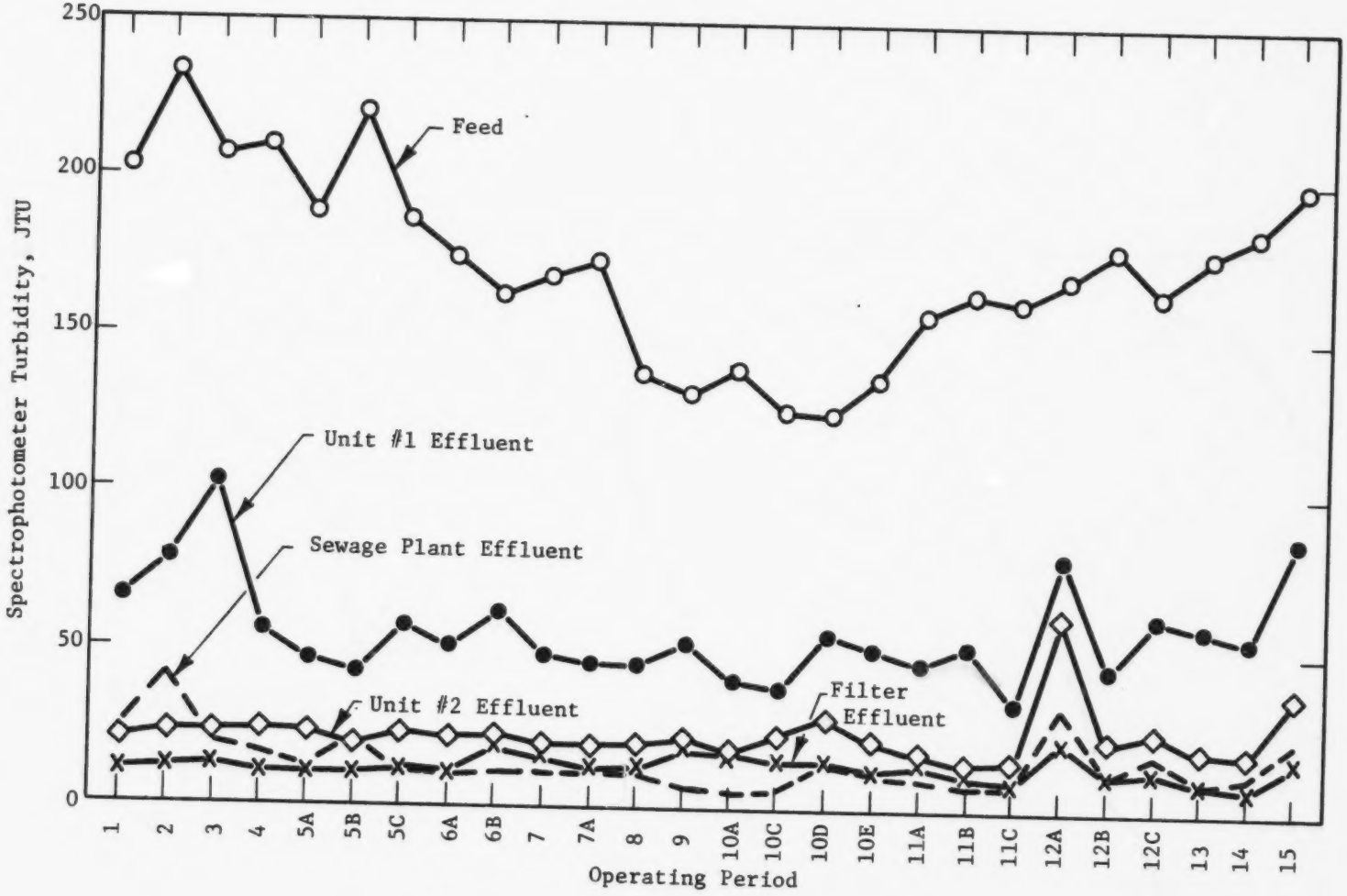

FIGURE 16: Mean Values of Spectrophotometer Turbidity

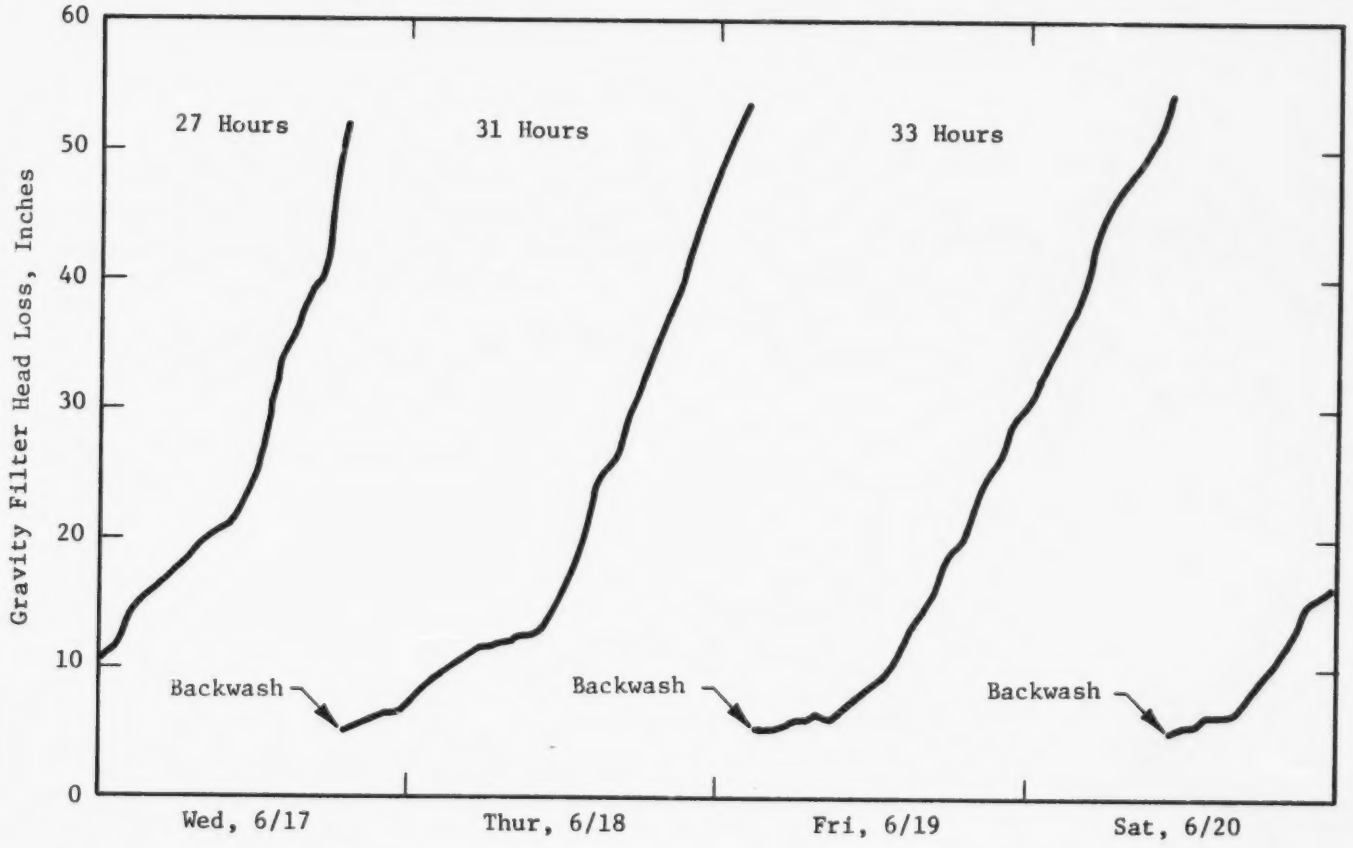

FIGURE 17: Typical Gravity Filter Head Loss Curves

Phase 3 - The pilot plant was operated for 10 days at conditions established in the previous period except Darco S-51 powdered activated carbon was substituted for Aqua Nuchar A. Operating Period 7 began by changing the carbon in the carbon feeder and allowing the existing carbon slurry to work out of the system in a normal fashion. As anticipated, the floc settleability changed radically. Mean values for blowdown given in Figure 13 do not reflect the progressive reduction in blowdown volume. At the end of Period 7 blowdown rates were 2.7 and 0.9 percent, respectively, for Units #1 and #2.

Product water quality was unchanged (Table 4) but the S-51 carbon altered the slurry produced in Unit #2. The slurry, as dewatered in the concentrator, became so dense it would not discharge consistently through the 1/2" blowdown valve and piping under the 5 ft of head available. Control of blowdown became erratic and by the middle of the third day the recirculating slurry solids had built up to 18 Vol %/5 min. Unit #2 was manually purged resulting in the loss of most of the slurry. Some blowdown must be maintained to sustain carbon advance so it required another three days, with a blowdown of 0.3% of throughput, to refill the concentrator and reestablish the recirculating slurry inventory. A 15-hour power failure precipitated another cycle of the blowdown problem so the run was terminated. The high solids blowdown is desirable, but it was not compatible with the scale of the pilot plant blowdown system.

Gravity filter runs averaged 35 hours for the period.

Operating Period 7A was a successful 12-day test using Darco S-51 and feeds of 6 mg/1 C-31/unit and 10 mg/1 alum to Unit #1 at the 1.5 gpm/ft^2 overflow rate. The concentrator in Unit #2 was bypassed by extending the internal piping to draw blowdown directly from the recirculating slurry. Bypassing the downstream concentrator required increasing the blowdown to 5.2 percent from the unit. Unit #1 blowdown was constant at 3.5 percent, the lowest mean rate recorded during the program except when treating less than 13.0 gpm (1.0 gpm/ft^2 overflow rate).

Filtrable-TOC and -COD removals were 59.5 and 68.0 percent, respectively, somewhat lower than for Period 7. Removals are higher if based upon Unit #2 effluent rather than filter effluent as filtrable-TOC and -COD levels increased measurably through the sand filter. This is especially evident after filter operation was changed in Period 5C. Determinations on filtered samples do not measure all of the adsorbate-indicator material going into the filter. A portion is particulate matter or is adsorbed upon the carbon carryover removed by the laboratory filtration. Some of this material appears to be solublized or desorbed in the filter and is measured in the filter effluent determinations. Filter runs averaged 34 hours for the period.

TABLE 4

Phase 3 & 4, Pilot Plant Performance by Operating Period, Mean Values

	Phase 3		Phase 4
	Operating Period		
	7	7A	8
CONTROLLED VARIABLES			
Overflow Rate, gpm/ft^2	1.5	1.5	1.5
Carbon Feed, mg/l	S-51,145	S-51,145	ANA,145
C-31 Feed, mg/l Unit #1	6	6	6
Unit #2	6	6	6
Alum Feed, mg/l Unit #1	10	10	10
Unit #2	0	0	0
No. of Days Sampled	7	10	12
No. of Operating Days	10	12	13
PILOT PLANT REMOVALS			
TOC unfiltered, %	90.7	89.0	90.3
TOC filtrable, %	67.7	59.5	70.7
COD unfiltered, %	87.7	87.4	87.1
COD filtrable, %	71.1	68.0	71.1
BOD_5 unfiltered, %	91	89	89
Sus. Solids, %	96.6	96.1	96.5
SEWAGE PLANT REMOVALS			
TOC unfiltered	88.8	87.8	87.5
COD unfiltered, %	88.8	89.0	88.7
BOD_5 unfiltered, %	92	90	93
Sus. Solids, %	97.0	97.6	95.7

The division in carbon loading between units was substantial during Period 7 when the carbon advance stream averaged 0.2 gpm. Loading was equal in the units during Period 7A with the carbon advance stream increased to 1.0 gpm. The volume of carrier liquid in the downstream blowdown constitutes a recirculating stream that dilutes the applied adsorbate concentration in the upstream contactor. A higher adsorbate concentration overflows to Unit #2. The greater the recycle the lower the loading potential becomes in Unit #1 and vice versa.

Phase 4 - The plant was operated for 13 days for Operating Period 8 repeating the conditions of the last three runs but again using Aqua Nuchar A as the adsorbent. The Unit #2 concentrator was returned to service.

In comparing the performance of the two carbons as given in Table 4, Aqua Nuchar A produced the higher filtrable-TOC and -COD removals, 70.7 and 71.1 percent, respectively. Examination of any of several figures, say Figure 11, however, reveals the performance differential was in the gravity filter. Filtrable-COD removal in the carbon contacting units, filter excluded, was 71.1 percent for both Periods 7A and 8. The same samples indicated filtrable-TOC removals of 65.7 and 70.4 percent, but these differentials are based upon analytical differentials approaching the limits of test accuracy.

Perhaps the 28-hour average filter runs for Period 8 contributed to improved filter performance. Wash water consumption was 5.0 percent primarily because operators did not work the wash cycle in a timely fashion.

Slurry blowdown rates with Aqua Nuchar A carbon were reestablished at 4.9 and 1.5 percent for Units #1 and #2, respectively.

As indicated in the earlier laboratory isotherm studies, there appeared to be little or no difference in adsorptive capacity between the carbons under the test conditions.

The relationship between the suspended solids content and the 5-minute settled volume of the recirculating slurries varies with the types and quantities of chemicals fed as well as with the character of the feed stream. The 5-minute settled volume test is an important clarifier operating parameter used to judge blowdown requirements. The suspended solids determinations can be used to estimate the carbon inventory in the system. Figures 18 and 19 show these relationships for slurries from the two clarifiers. The curves on these figures illustrate the effect of different chemical feed situations and indirectly imply that a wide range in possible carbon inventory was experienced during the program. In general, Unit #1 slurry was composed of about 2 parts raw sewage solids to 1 part of carbon based upon feed solids and carbon feed figures. This minimized the effect of carbon type in Unit #1 (Figure 18)

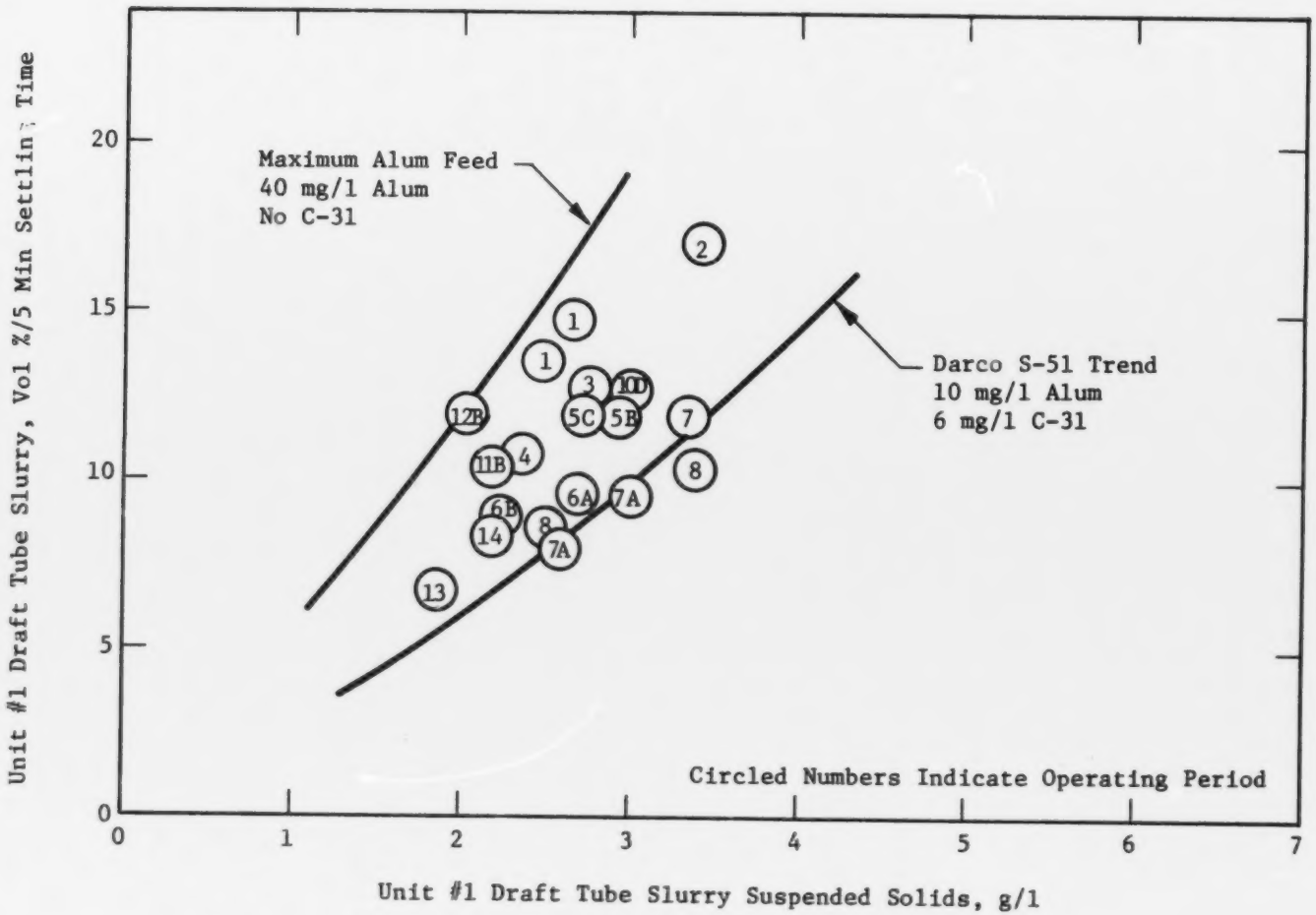

FIGURE 18: Solids Content of Recirculating Slurry, Unit #1

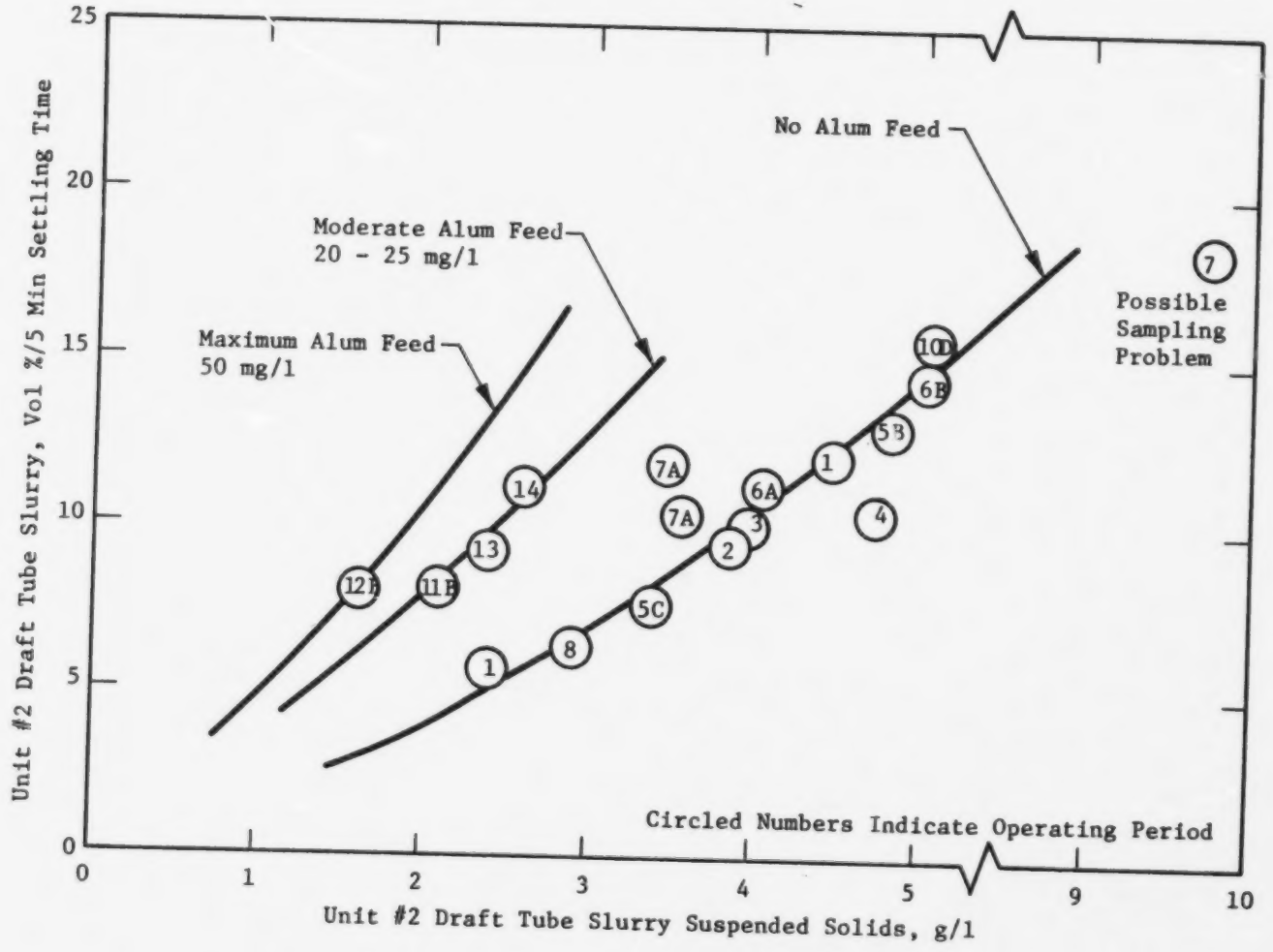

FIGURE 19: Solids Content of Recirculating Slurry, Unit #2

as indicated by the points for Periods 7A (S-51) and 8 (ANA). In Unit #2 (Figure 19), the Period 7A points stand apart from the others, possibly resulting from the atypical blowdown situation for this period only. This altered the dilution and recirculation ratios existing in the mixing and flocculation zones of the unit.

For much of the program the activated carbon inventory was estimated to be 7 to 9 lb with the countercurrent flow-through time being about 6 hours while operating at 1.5 gpm/ft^2. Although 5-minute settling test values averaged about 10 percent volume in both units and both contained about 225 gal. of slurry, the carbon was not divided evenly between units. Unit #1 slurry had a lower suspended solids content, say 2.5 g/l, and only 37 percent of it could be considered carbon. In Unit #2 with about 4 g/l suspended solids content, close to 78 percent of it could be carbon. Including concentrator contents Unit #1 probably operated with 2 lb and Unit #2 with 6 lb of the inventory.

Settling rate curves (Figure 15) were not too helpful in discriminating between the ANA and S-51 carbons. Both slurries settled so rapidly that points could seldom be obtained before hindered settling commenced at about 1.5 to 2.0 minutes into the test. The settling solids set up density currents which, along with residual currents from the initial sample mixing, obscured the liquid-solids interface during the crucial 1st minute of the test. In the 250-ml graduated cylinder (10.25 in. calibrated depth) used for the tests, Unit #1 slurry was estimated to have a free subsidence rate in excess of 4 in./min or greater than 2.5 gpm/ft^2 overflow rate. Unit #2 slurry was estimated to settle at twice that velocity.

Phase 5 - The program proceeded with 53 days of operation divided into 9 separate Operating Periods, each utilizing a variation in coagulant dosage. Operating Period 10B was a 1-day transition period and data for that day were discarded. Operational information for the other 8 periods is presented in Table 5.

Period 9 was a test of operation using 4 mg/l C-31 per unit to establish the minimum dose for effective flocculation of the carbon. During Period 9, the suspended solids carry-over from both units nearly doubled and the amount of uncoagulated raw-sewage colloidal material passing the filter more than tripled the values reported during the preceding work with 6 mg/l C-31 per unit. This is shown clearly in Figure 8. Other parameters were less affected as adsorption undoubtedly continued at established efficiency whereas analytical determinations involving 0.45-micron filtration reflect the passage of uncoagulated particulate matter. Lab-filtered samples contained discernible turbidity.

TABLE 5

Phase 5, Pilot Plant Performance by Operating Period, Mean Values

	Operating Period							
	9	10A	10C	10D	10E	11A	11B	11C
CONTROLLED VARIABLES								
Overflow Rate, gpm/ft^2	1.5	1.0	1.0	1.0	1.0	1.0	1.0	1.0
Carbon Feed, mg/l ANA	145	145	145	145	145	145	145	145
C-31 Feed, mg/l Unit #1	4	4	4	4	6	6	6	6
Unit #2	4	4	6	6	6	6	6	6
Alum Feed, mg/l Unit #1	10	20	20	25	0	10	10	15
Unit #2	0	0	0	0	25	25	20	20
No. of Days Sampled	5	6	4	7	5	6	6	6
No. of Operating Days	5	6	5	8	7	7	7	7
PILOT PLANT REMOVALS								
TOC unfiltered, %	87.4	90.0	88.8	88.8	89.5	91.3	92.1	92.2
TOC filtrable, %	60.6	66.7	60.8	52.3	67.3	73.6	71.1	70.1
COD unfiltered, %	83.0	85.3	83.2	81.8	86.7	90.2	90.0	90.6
COD filtrable, %	59.7	69.3	64.4	54.3	65.5	70.9	69.0	70.2
BOD_5 unfiltered, %	-	80	89	83	90	95	93	91
Sus. Solids, %	88.5	95.1	93.3	92.7	96.6	97.1	96.0	98.8
SEWAGE PLANT REMOVALS								
TOC unfiltered, %	89.6	90.8	90.1	86.3	87.4	88.2	89.0	90.2
COD unfiltered, %	90.6	90.8	88.1	84.7	86.5	88.9	89.5	89.5
BOD_5 unfiltered, %	-	86	87	93	91	91	90	86
Sus. Solids, %	93.9	97.9	97.5	93.7	97.5	97.1	97.4	98.3

1 day period 10B was transitional and was discarded

The overflow rate was decreased to 1.0 gpm/ft^2 for the balance of the program at this time. This reduced the rate of chemical consumption, eliminated nuisances associated with hydraulics at the no-safety-factor plant design capacity and brought solids production more in line with slurry concentrator capacity.

Filter alum precipitates 0.263 parts of inert solids (aluminum hydroxide) for each part fed. If alum equal to 10 percent of the carbon dose was fed to a system utilizing carbon regeneration and reuse with a 15 percent carbon loss per cycle, the inert-solids fraction of the mixture would build up to 14.9 percent. (If the 15 percent loss is defined as solids mixture rather than only carbon, the inert-solids fraction would approach 17.1 percent.) Should the inert aluminum hydroxide be detrimental to regeneration its use might better be avoided. This would be true only if the alum interference was specific because the screened and degritted sewage feed stream was found to contain 65, 52 and 80 mg/l fixed suspended solids on composite samples having 204, 204 and 272 mg/l total suspended matter, respectively. These quantities, with the 145 mg/l carbon feed of the pilot plant, produce a mixture that would be 31.0, 26.4 and 35.6 percent inert solids in one pass through the plant. The slurry for a regeneration-reuse system with 15 percent carbon loss per cycle would equilibrate at 75.0 percent inert material for the 65-mg/l sample. A modest alum feed would not be the predominant source of inert solids in such a system. A decision was made to study higher alum dosages supplanting portions of the polymer dosage.

For Operating Period 10A, the alum dosage to Unit #1 was increased from 10 mg/l to 20 mg/l with other feeds held constant. Filtrable-TOC and -COD removals improved as the filtrable suspended matter was reduced. Suspended solids carryover decreased to acceptable levels for both units but solids (12 mg/l) and turbidity (18 JTU) passing the gravity filter remained undesirably high (Figures 8 and 16). The slurry in Unit #2 was gradually deteriorating so the polymer feed was increased from 4 mg/l to 5 mg/l (Period 10B). After 24 hours little improvement was noted so the polymer feed was increased again to 6 mg/l (Period 10C). This was a time (August) of weak raw sewage and although pilot plant effluent characteristics were little changed, percent removals decreased. Carbon loading in the clarifiers dropped to 8.4 percent as TOC and 31.0 percent as COD.

Blowdown was 4.2 percent for Unit #1 and 0.8 percent for Unit #2 reflecting the influence of the reduced treating rate.

Gravity filter runs averaged 34 hours during Period 10A prior to floc deterioration then deteriorated to 20 hours. They lengthened to 29 hours for Period 10C. Wash water consumption was 4.3 percent with a filter surface loading of 2.75 gpm/ft^2 for this later period.

The alum dosage was set at 25 mg/1 to Unit #1 for Period 10D. Operation continued at a submarginal condition with Unit #1 carryover increasing beyond the 50 mg/1 suspended solids limit. Solids and turbidity passing the filter remained undesirably high. Carbon loading, as a result of low feed stream strength and filtrable solids problems was 7.0 percent as filtrable TOC and 28.1 percent as filtrable COD, the lowest encountered at this carbon dosage.

Blowdown and gravity filter operation remained essentially unchanged from the previous period.

The 25 mg/1 alum feed was switched to Unit #2 and the polymer feeds reset to 6 mg/1 per unit for the 7-day Operating Period 10E. All removal figures improved considerably as filter effluent suspended solids and turbidity decreased to 7 mg/1 and 13 JTU, respectively, and feed stream strength began recovering.

Carbon loadings in the combined clarifiers went to 9.5 percent as filtrable TOC and 31.5 percent as filtrable COD.

Blowdown requirements shifted substantially to 6.1 percent for Unit #1 and 2.7 percent for Unit #2 with its alum-floc load now imposed upon both clarifiers.

The 20 mg/1 of alum-coagulated solids carryover to the gravity filter was within limits but the filter runs decreased to 19 hours. With the shift in flocculation potential the slurry inventory in Unit #1 remained more uniform. Solids escaping the 6 mg/1 polymer in this unit during peak solids influx periods of the day were captured in Unit #2 causing its slurry inventory to cycle diurnally.

The phase was completed with 3 periods, 11A, 11B and 11C, of one-week duration each, studying the use of alum in both clarifiers. Table 5 and the several figures show this to be a highly satisfactory operation with all removals as high or higher than for any previous period. Feed stream strength was increasing.

Filtrable-TOC and -COD loadings in the clarifiers increased to 11.4 and 37.2 percent, respectively.

The greater total alum dosage (30 to 35 mg/1) doubled the blowdown rate to 12.4 percent for Unit #1. Unit #2 blowdown decreased to 1.5 percent from a high of 2.8 percent as a result of the greater solids capture in Unit #1.

Gravity filter operation improved. Filter runs lengthened to 36-hour average duration and wash water consumption decreased to 2.5 percent of the volume filtered.

The slurry solids inventory relationships for part of Period 11A are reflected in the field record of draft tube slurry volume tests, Figure 20. During the low solids influx period of Saturday and Sunday, the inventory in the units cycled about equally; however, by Sunday evening the increasing load plus changes in the character of the solids reached a state wherein the potential for solids capture, i.e., the coagulation system capacity, for Unit #1 was exceeded. The inventory equilibrated. Uncaptured solids passed on to Unit #2 causing wide diurnal cycles Monday and Tuesday in spite of numerous blowdown timer changes to limit the extremes.

Phase 6 - The pilot plant was operated for 29 days with alum as the primary coagulant. Dosages of 20 mg/1 per unit were inadequate so they were increased stepwise per Table 6 to 40 mg/1 to Unit #1 and 50 mg/1 to Unit #2 followed by readjustment finally to 50 mg/1 to Unit #1 and 25 mg/1 to Unit #2.

Performance for the initial two days was acceptable; however, these data were discarded as the existing polymer-bearing floc was being purged from the system. As this occurred, uncoagulated carbon began discharging in the clarifier effluents. Filter runs became as short as 3 hours, with carbon penetrating the media. Filter operation was suspended intermittently while alum feeds were increased and, ultimately, polymer was added to restore floc condition.

The plant could not be operated successfully at the 1.0 gpm/ft^2 overflow rate without polymer. The bulky alum floc overwhelmed blowdown capabilities, filled the clarifiers to the launder elevation and decanted downstream in large amounts. Polymer restored slurry control to a degree allowing some reduction in blowdown.

The concentrator in Unit #1 was enlarged to 15 gal. capacity for the final operating period of this phase.

The data are shown in the several Figures; however, they are presented in a qualified sense that not enough determinations could be retained to present a representative picture. In addition, the excessive and variable blowdown altered the carbon/feed-stream ratio beyond reasonable reconciliation. This later condition is believed responsible for the higher indicated removals in this phase.

Phase 7 - The pilot plant had performed through most of the previous operating periods with 145 mg/1 activated carbon feed at efficiencies closely approximating those for the sewage treatment plant. Figures 21, 22, 23 and 24 suggest, aside from problem periods, that both systems were governed by a common denominator despite their process differences.

The pilot plant was operated at increased carbon dosages of 200 mg/1 for 21 days and 250 mg/1 for 13 days comprising Operating Periods 13 and 14 during October and November. This was followed 11 weeks later

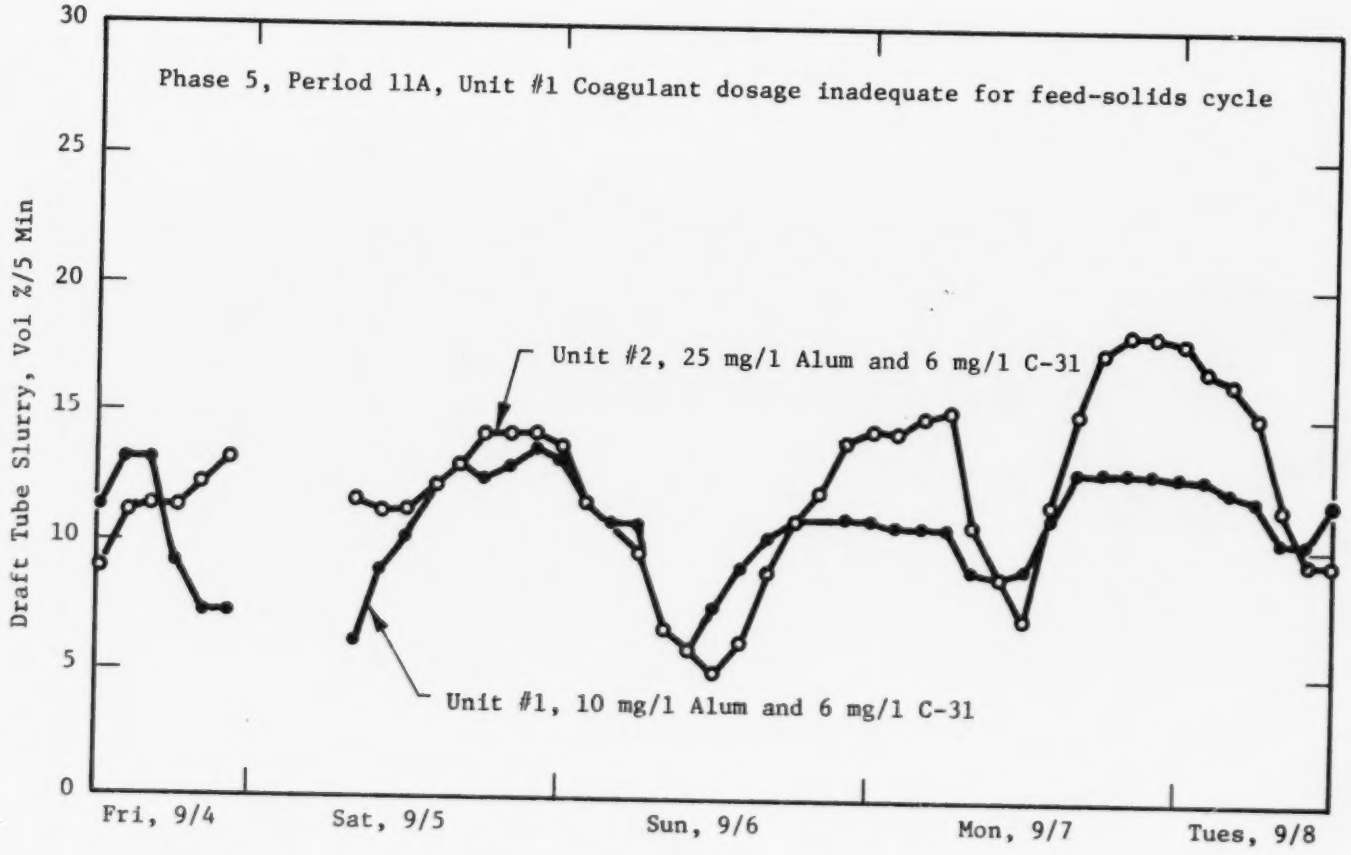

FIGURE 20: Diurnal Variation, Clarifier Slurry Tests, Phase 5

TABLE 6

Phase 6, Pilot Plant Performance by Operating Period, Mean Values

	Operating Period		
	12A	12B	12C
CONTROLLED VARIABLES			
Overflow Rate, gpm/ft^2	1.0	1.0	1.0
Carbon Feed, mg/l ANA	145	145	145
C-31 Feed, mg/l Unit #1	0	0	0-1-2
Unit #2	0	0-3-0-3	3-2-3
Alum Feed, mg/l Unit #1	20-30-40	40	40-50
Unit #2	20-30-40-50	50	50-25
No. of Days Sampled	2	3	5
No. of Operating Days	7	7	15
PILOT PLANT REMOVALS			
TOC unfiltered, %	93.6	94.0	93.6
TOC filtrable, %	76.9	79.4	80.2
COD unfiltered, %	90.7	91.7	90.2
COD filtrable, %	70.2	80.3	78.1
BOD_5 unfiltered, %	88	90	96
Sus. Solids, %	93.5	98.7	97.0
SEWAGE PLANT REMOVALS			
TOC unfiltered, %	88.4	87.0	85.3
COD unfiltered, %	87.8	85.7	84.0
BOD_5 unfiltered, %	65	87	94
Sus. Solids, %	93.0	97.9	89.3

Hyphenated chemical dosage data indicate changes instituted during period

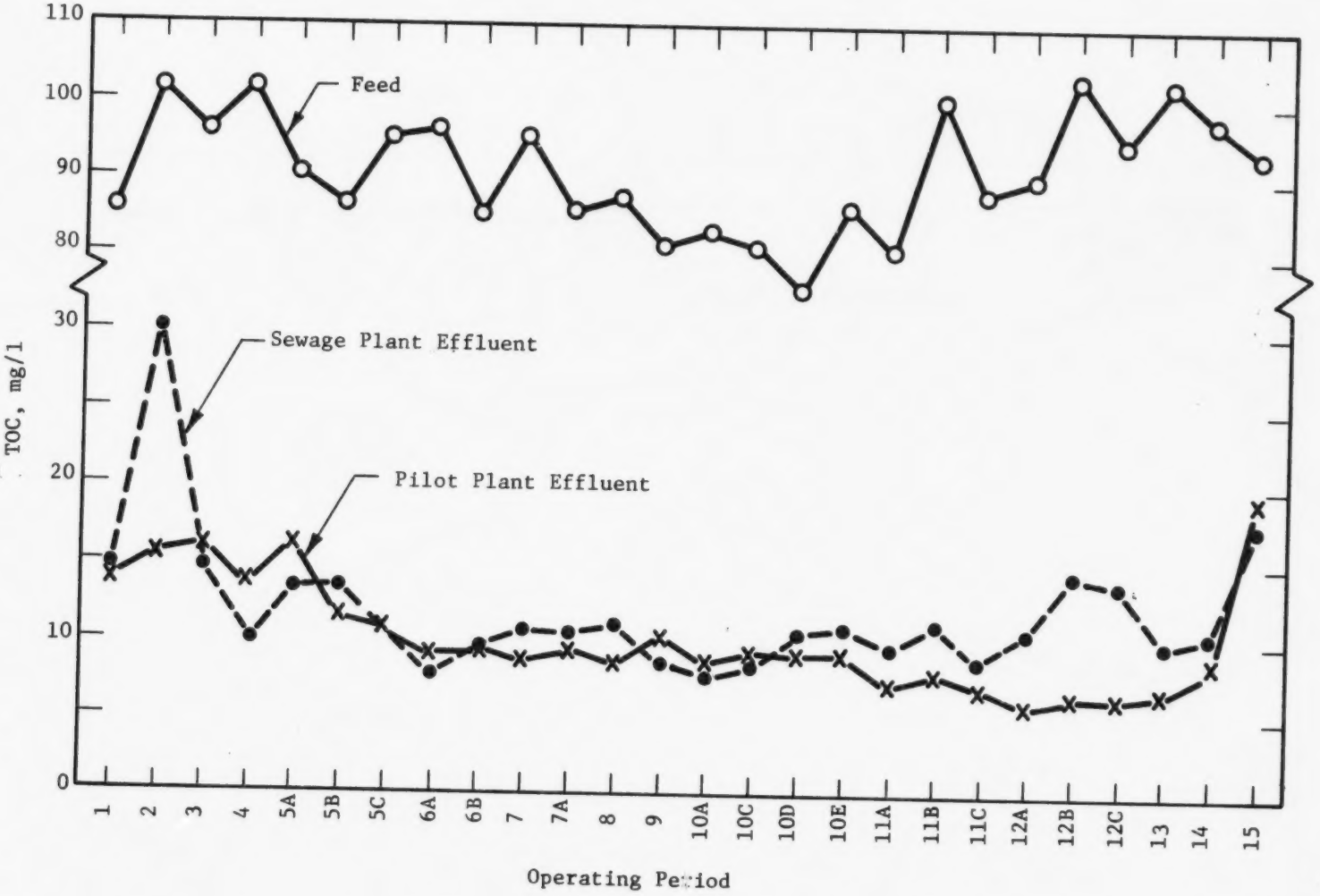

FIGURE 21: Mean Values of TOC

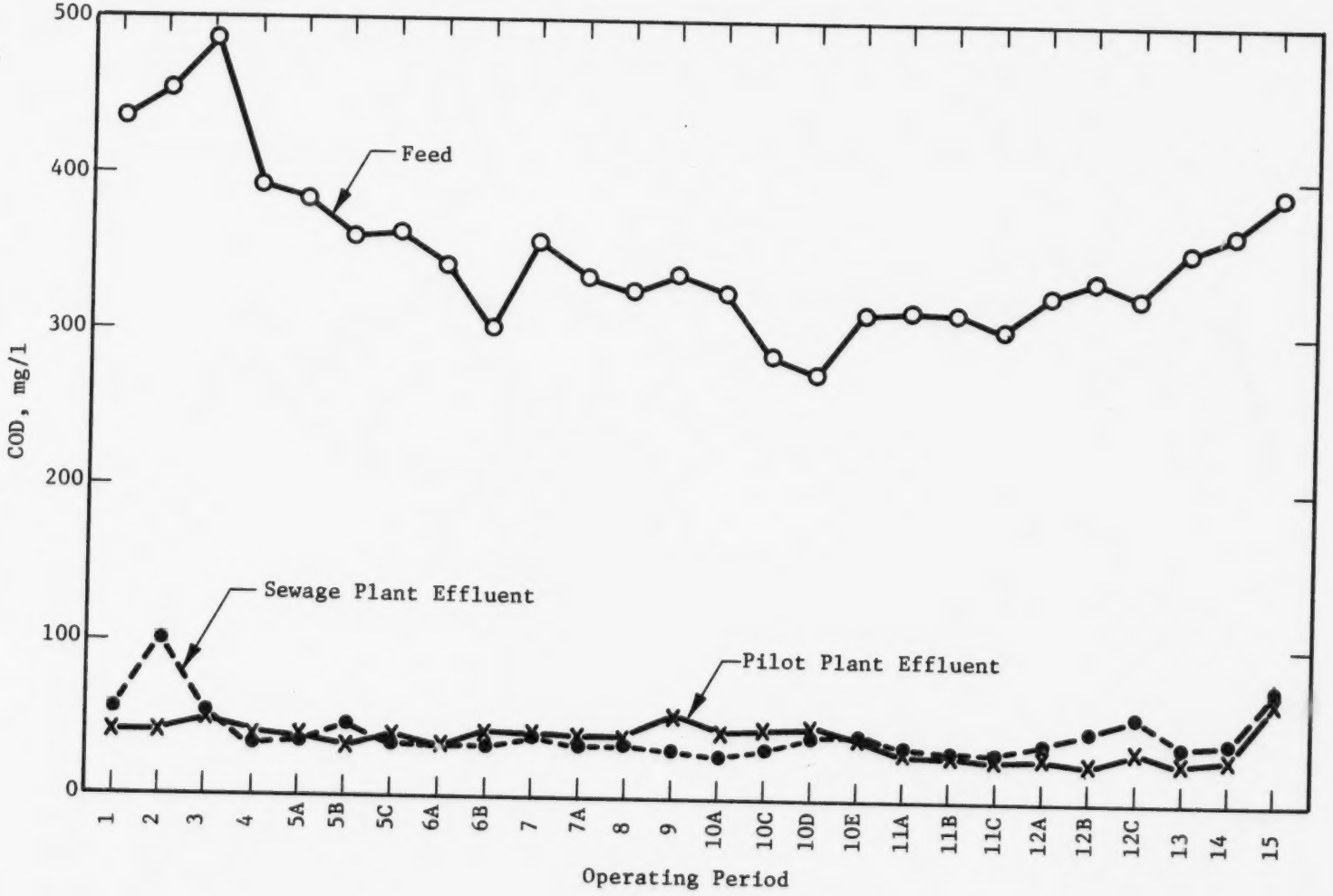

FIGURE 22: Mean Values of COD

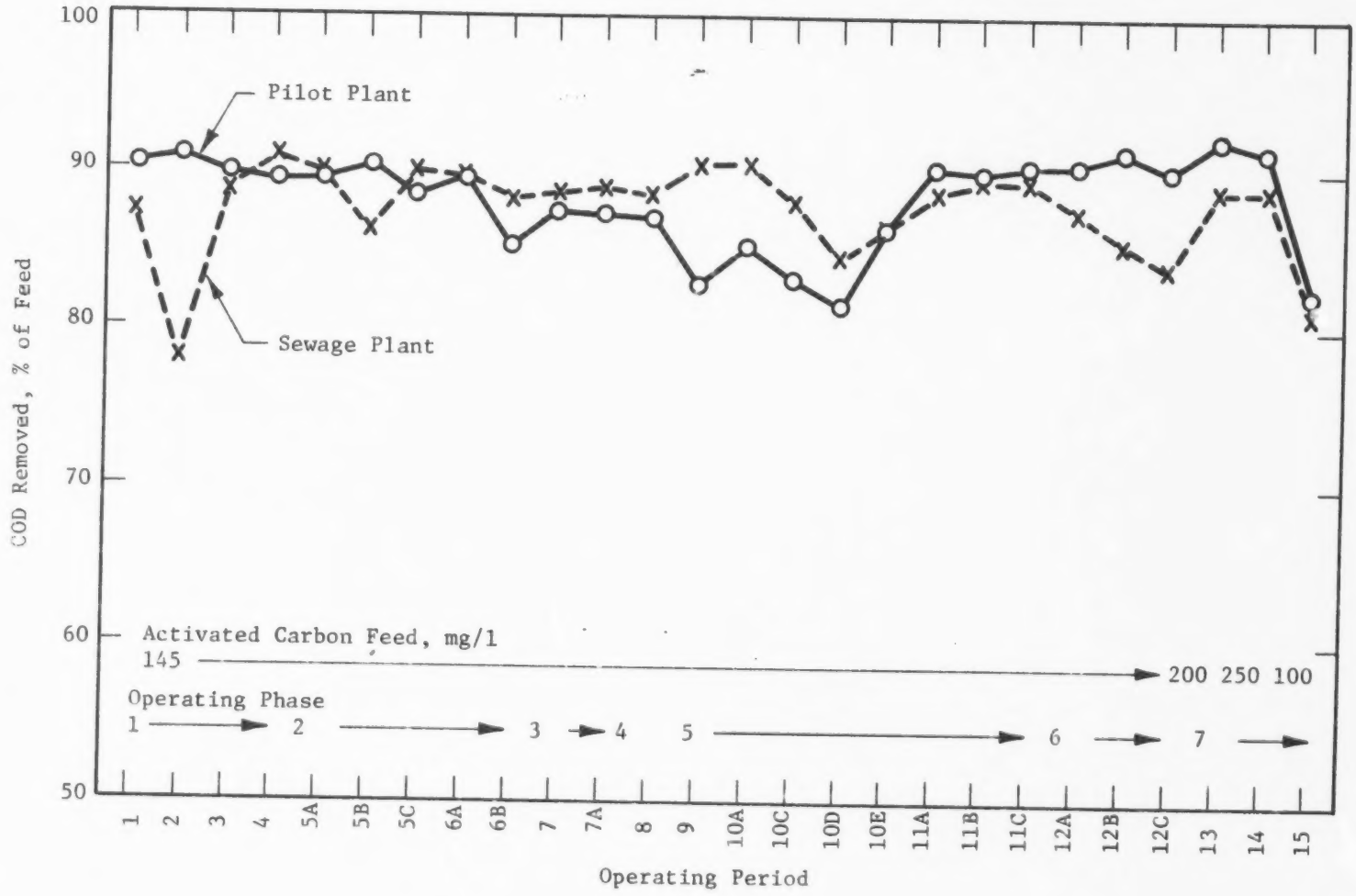

FIGURE 23: Mean Values of COD Removal

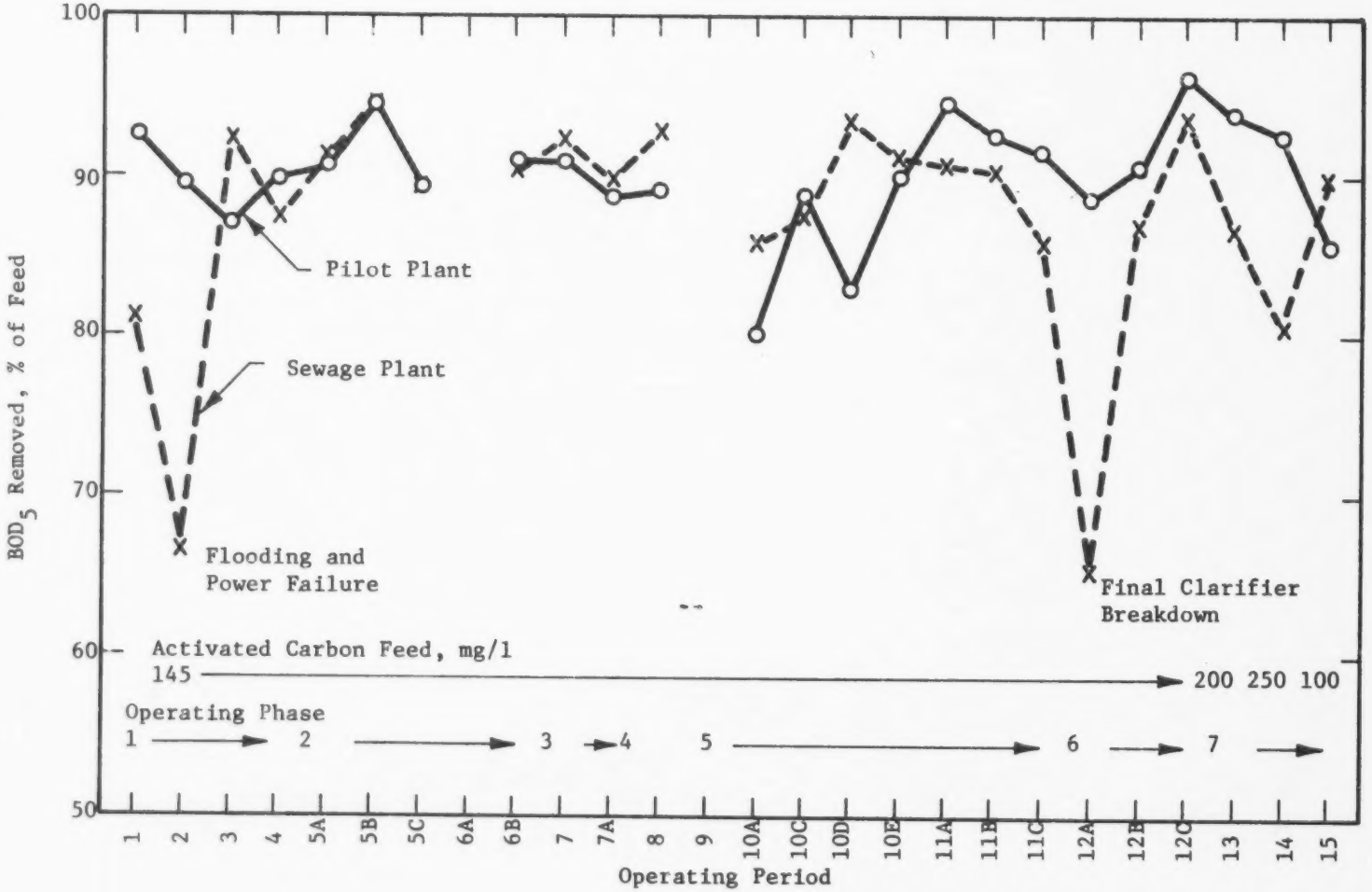

FIGURE 24: Mean Values of BOD_5 Removal

by the final 13-day Period 15 with 100 mg/l carbon. The objective was to deviate from the performance level of the activated sludge plant by exploring the limits of treatability and control of treatment.

Table 7 shows all operations were conducted at moderate levels of polymer and alum dosages. The only data discarded were for transitional days and for restarting the plant from a drained condition in the case of Period 15.

Removal of filtrable TOC and COD was 78.0 and 78.2 percent, respectively, for Period 13 and 73.7 and 74.6 percent, respectively, for Period 14. These values plus those for total TOC and COD removal were the highest obtained for the field program (excepting qualified Phase 6). The feed stream TOC and COD levels were increased during this portion of the year but not high enough to offset the increased carbon dosage effect. The process did not respond in accord with isotherm formulae but was limited by the fraction of TOC- and COD-exhibiting materials not readily adsorbed.

Activated carbon loadings for Periods 13 and 14 averaged 9.7 and 8.0 percent as filtrable TOC with applied loads of 13.0 and 10.3 percent, respectively. The applied loading, i.e., the feed stream filtrable TOC expressed as a weight percentage of the carbon dose, having decreased as a result of the higher carbon feeds. For the same periods, loadings of 36.8 and 30.1 percent as filtrable COD resulted from applied loads of 47.3 and 38.5 percent, respectively. All loading data refer only to the clarifiers, i.e., gravity filter influence is excluded.

Clarifier blowdown (Figure 13) reflected the use of alum in both clarifiers, the higher solids load contributed by the feed stream and carbon feeds and, during Period 13, the increase in Unit #1 concentrator volume. This later effect was counteracted during Period 14 by still higher feed solids and carbon feed.

Filter runs averaged 29 hours and 25 hours for the two periods with wash water consumption reported as 3.2 and 3.8 percent of the filtered volume. Examination of accumulated data implied pilot plant carbon dosage exceeded that required to place performance of the adsorption system in the linear portion of the isotherm curve. Carbon dosage for Period 15 was selected to be equal to the applied filtrable COD, i.e., 100 mg/l estimated; such that a carbon loading of 70 percent as filtrable COD would have to be achieved to approximate the degree of removal reported for many of the previous periods. Loadings of this amplitude had been indicated in two grab sample isotherms but only at applied COD/carbon ratios exceeding 2.5 and were never achieved with composite sample isotherms.

Mean applied filtrable COD was 102.8 mg/l for Period 15, which was scheduled in February. Filtrable-TOC and -COD removals were 47.0 and

TABLE 7

Phase 7, Pilot Plant Performance by Operating Period, Mean Values

	Operating Period		
	13	14	15
CONTROLLED VARIABLES			
Overflow Rate, gpm/ft^2	1.0	1.0	1.0
Carbon Feed, mg/l ANA	200	250	100
C-31 Feed, mg/l Unit #1	6	6	6-8
Unit #2	6	6-8	6
Alum Feed, mg/l Unit #1	10	10	15
Unit #2	25	25	15-20
No. of Days Sampled	19	12	10
No. of Operating Days	21	13	13
PILOT PLANT REMOVALS			
TOC unfiltered, %	93.3	91.5	79.5
TOC filtrable, %	78.0	73.7	47.0
COD unfiltered, %	92.2	91.6	82.4
COD filtrable, %	78.2	74.6	47.8
BOD_5 unfiltered, %	94	93	86
Sus. Solids, %	96.8	98.2	95.3
SEWAGE PLANT REMOVALS			
TOC unfiltered, %	90.3	89.6	81.4
COD unfiltered, %	89.1	89.0	81.0
BOD_5 unfiltered, %	87	80	90
Sus. Solids, %	95.3	96.1	89.8

Hyphenated chemical dosage data indicate changes instituted during period

47.8 percent, respectively. These data increased to 50.3 and 49.8 percent for the contactor-clarifier portion of the system only. Carbon loadings, also excluding the filter, were 15.9 percent as TOC and 51.1 percent as COD.

Total COD and TOC removal Figures 23 and 25 show sewage plant performance was reduced by nearly the same amount as the pilot plant for this time. Mean data for the sewage plant, however, fail to show that a process upset was discharging activated sludge in the effluent early in the period. This was followed by a recovery trend during the final 7 sampling days. Figure 26 shows this trend and, that prior to shutdown, the pilot plant effluent contained 14 mg/l more COD than that of the sewage plant. The magnitude of the loss in COD removal resulting from the reduced carbon feed was 20.7 mg/l COD. Color removal data (Figure 27) show a marked decrease in color reduction for the pilot plant as compared with the sewage plant. Color was the only solids-free adsorbate indicator monitored for both plants.

Suspended solids in the feed were high for the period and the preselected coagulation system was marginal. Solids carryover from both clarifiers exceeded the arbitrary limits mentioned earlier. Both polymer and alum feed were adjusted upwards during the run. Blowdown requirements increased to 7.5 percent for Unit #1 and to 5 percent for Unit #2.

The gravity sand-coal filter media had been removed, inspected and returned to the filter prior to operation. No evidence of channeling, mud balls or unexpected biological activity was discovered. With suspended solids carryover of 25 to 45 mg/l going onto the filter the 22-hour filter runs are reasonable. Wash water consumption was 3.8 percent of the volume filtered.

Supplemental Investigations

Additional data collected during the pilot plant program included the adsorption isotherm data, limited temperature measurements, pH monitoring, blowdown slurry solids content, color monitoring, limited nutrient measurements, turbidity measurement with comparisons and gravity filter operation.

Adsorption Isotherms - Adsorption isotherms produced from 24-hour composite samples during the field program were to have been the principal tool in evaluating process efficiency. Because of the successful use of isotherms in earlier programs involving tertiary treatment, it was unexpected when necessary laboratory procedures introduced factors negating the relationships between isotherm and plant data.

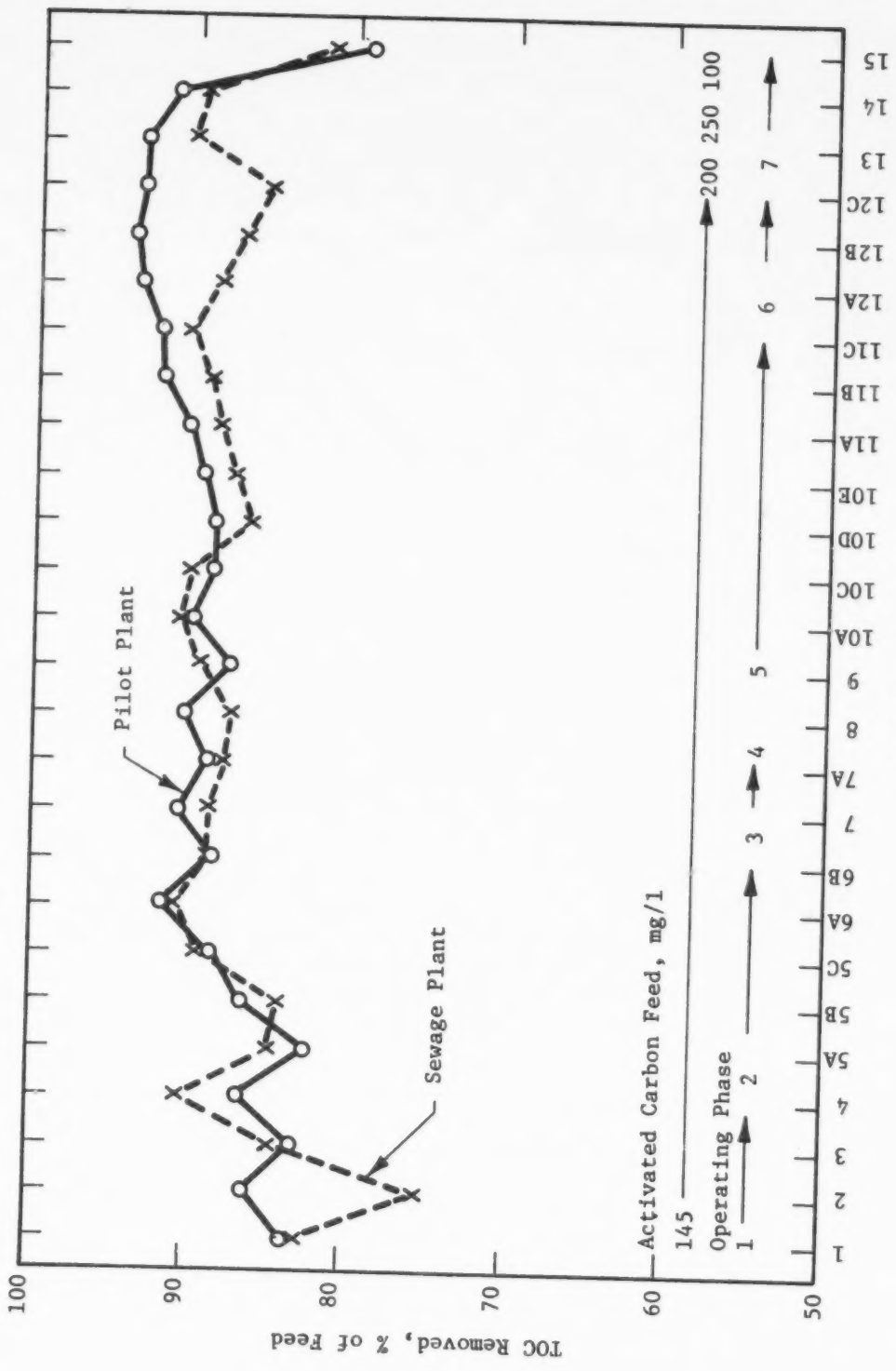

FIGURE 25: Mean Values of TOC Removal

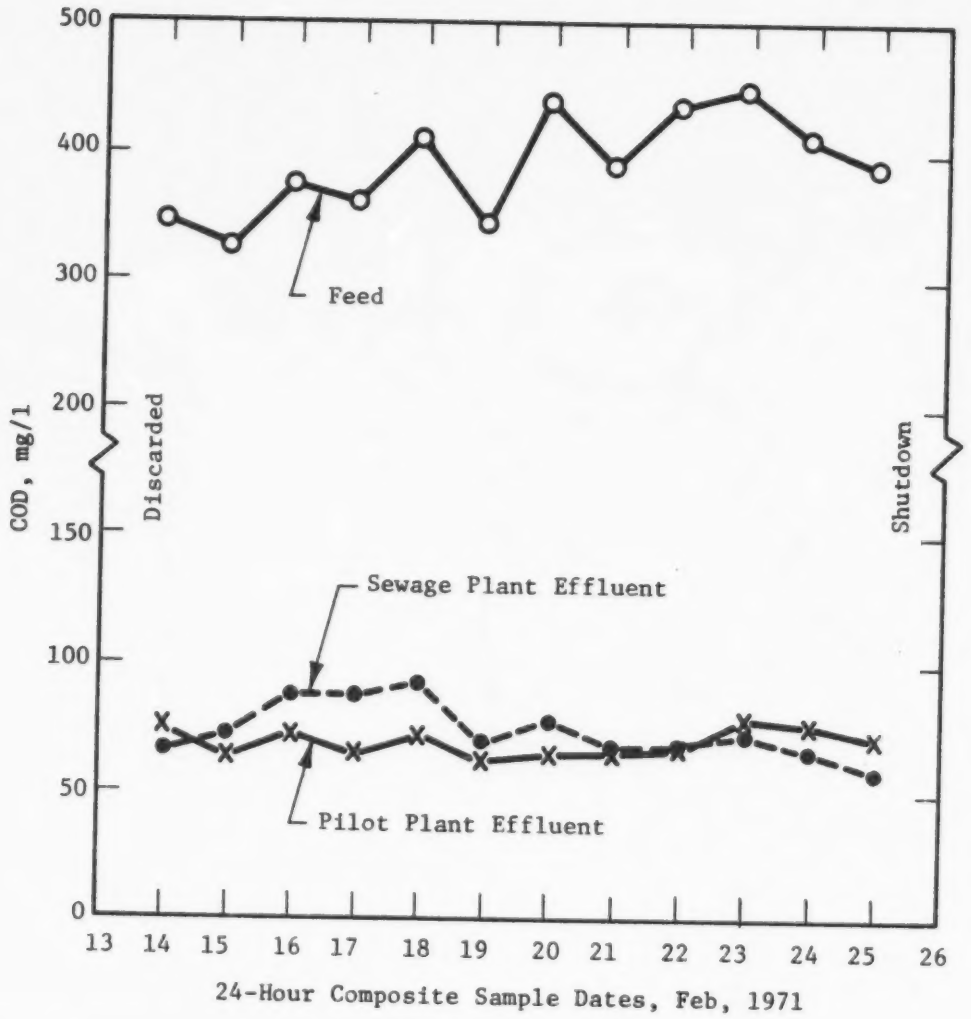

FIGURE 26: Daily Variation in COD, Period 15

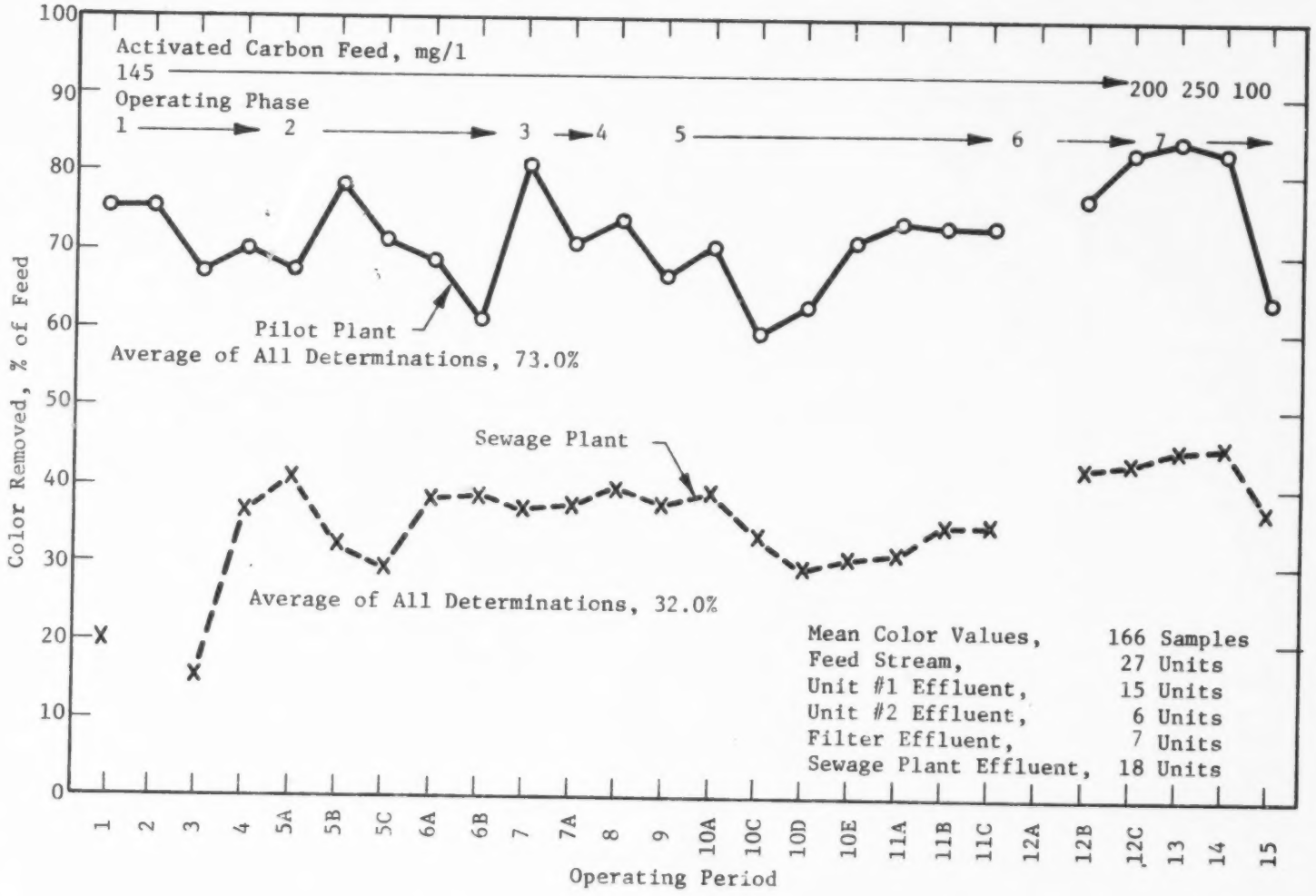

FIGURE 27: Mean Values of Color Removal

The difficulties were related to filtrable suspended solids and volatile components in the samples. To produce an adsorption isotherm, a unit influent composite sample was vacuum filtered through a 0.45-micron filter. A portion of the filtrate was analyzed directly producing the 24-hour sampling period results. The balance was used in the set of six jars for the isotherm. The isotherm blank was always treated with the coagulant dosage in use in the pilot plant unit during the sampling period and it underwent a final vacuum filtration to duplicate the handling of the carbon bearing jars. The initial filtrate usually had a higher organic carbon content and COD than the resulting isotherm-blank filtrate. Some organic filtrable suspended solids of less than 0.45 micron apparently were flocculated and retained on the second filter. The additional vacuum application possibly removed volatile material that otherwise might exhibit COD or contain organic carbon. The resulting adsorption isotherm, while perfectly acceptable when taken by itself, represented a system operating on a lower C_o datum than the pilot plant with insufficient information available to adjust for the differential.

The differences generally were significant. For example, a feed stream composite was found to contain 114.0 mg/l filtrable COD with the resulting isotherm blank testing 81.5 mg/l. Another tested 109.2 mg/l and 63.6 mg/l, respectively. Unit #2-influent isotherms were in better agreement with 60.0 mg/l and 59.1 mg/l reported for the first example above and 22.5 mg/l vs. 17.5 mg/l for the second. Better agreement would be expected because Unit #2 influent had been subjected to flocculation in Unit #1, 20 mg/l polymer in the first example and 6 mg/l polymer and 10 mg/l alum in the second.

The first Unit #2-influent isotherm example given above is one of the few considered suitable for evaluating the pilot plant operation. It is reproduced in Figure 28 along with actual pilot plant operating points for the same date. The points lie on curves labeled performance paths which represent the variation in carbon loading across a range of C_t values while the carbon dosage of 145 mg/l and the C_o values are constant.

The isotherm represents single-stage adsorption performance with carbon dosage varied so that the intercept of the Unit #2 performance path and the isotherm, at a loading of 12.1 percent and C_t = 42.3 mg/l COD, establishes single-stage adsorption efficiency for 145 mg/l carbon. Unit #2, however, demonstrated process superiority by producing a loading of 19.3 percent at C_t = 32.0 mg/l COD. COD removal in Unit #2 was 47 percent, i.e., reduced from 60.0 mg/l to 32.0 mg/l. The single-stage carbon loading indicated by the isotherm at C_t = 32.0 mg/l is 8.0 percent establishing the single-stage carbon requirement, to reach a C_t of 32.0 mg/l, to be 350 mg/l.

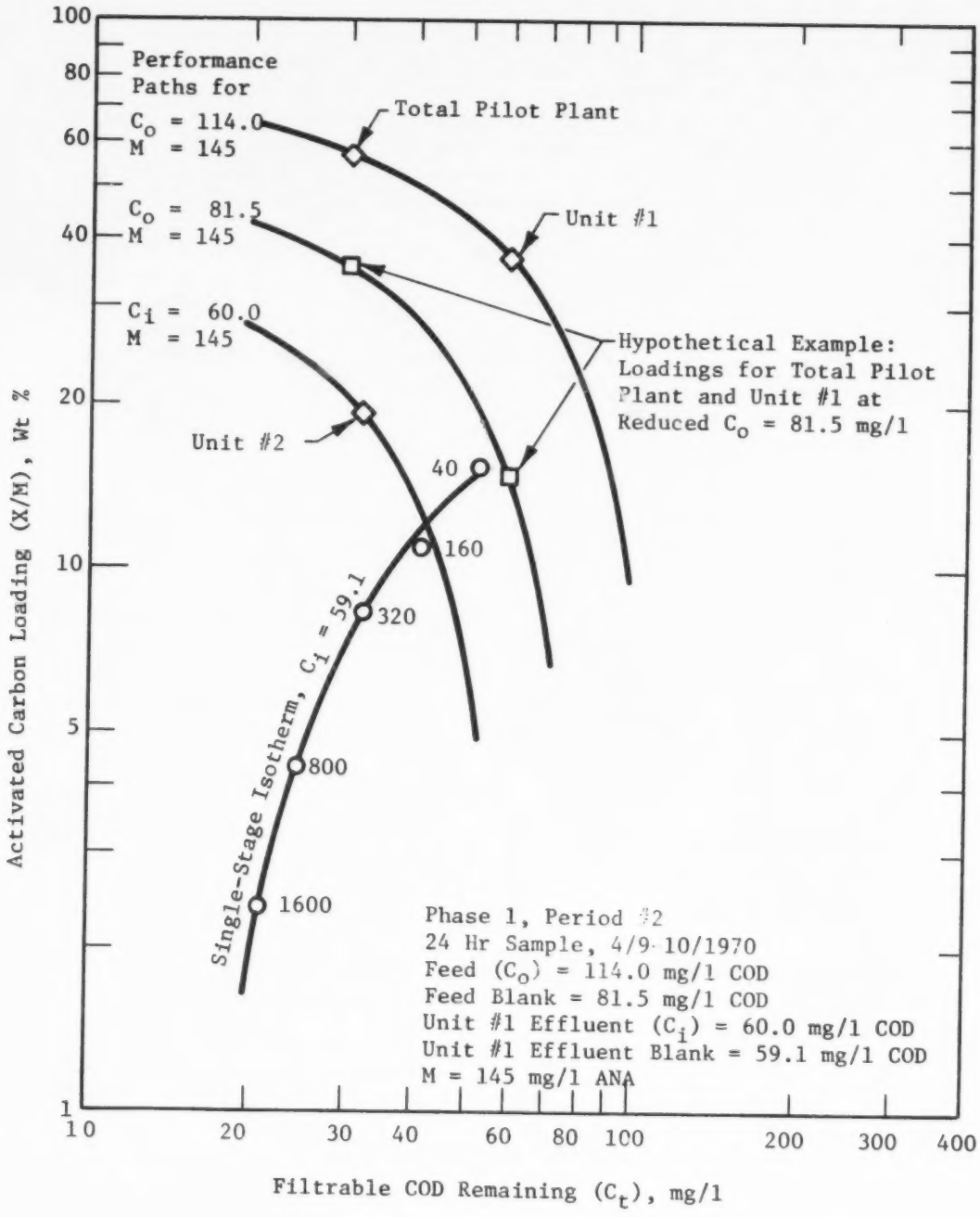

FIGURE 28: Freundlich Adsorption Isotherm Pilot Plant Analysis

Published information for predicting two-stage countercurrent systems did not cover the entire range of adsorbate-residual values of interest so available data were extrapolated graphically. Assuming the principal slope of the isotherm (1/n = 2.3) to apply, two-stage countercurrent adsorption requires 42 percent of the 350 mg/l single-stage carbon dose for the removal achieved. This is 147 mg/l carbon or essentially that actually used in Unit #2. As suggested in the original work[5], the single contactor-clarifier was producing two-stage countercurrent adsorption performance.

In the pilot plant the once-used carbon was then advanced to Unit #1 where an additional loading of 37.2 percent was achieved. In so doing, the feed COD of 114.0 mg/l was reduced to the intermediate concentration (C_i) of 60.0 mg/l. On the downstream end of the pilot plant the gravity filter removed an additional 1.6 mg/l COD delivering a final effluent of 30.4 mg/l filtrable COD. The total plant removed 83.6 mg/l filtrable COD resulting in a carbon loading of 57.6 percent and 73.3 percent filtrable COD removal.

If the reduction in filtrable COD between the feed stream determination of 114.0 mg/l and the Unit #1 influent isotherm blank determination of 81.5 mg/l is considered physical removal of suspended matter, a hypothetical correction in loadings can be made. A hypothetical performance path for C_o = 81.5 mg/l COD and 145 mg/l carbon is included on Figure 28. The total pilot plant removal becomes 51.1 mg/l COD with a total loading of 35.2 percent. Unit #1 loading is reduced to 14.8 percent; however, this is still load added to the once-used carbon which brought an unchanged 19.3 percent load with it from Unit #2. By comparison, the Unit #1 influent isotherm (not shown) prepared with virgin carbon and with C_o = 81.5 mg/l COD gave the loading for 145 mg/l carbon as 17.5 percent at C_t = 56.0 mg/l. The curve broke sharply in this region never exceeding 18.4 percent loading.

In the hypothetical analysis, adsorbate removal was 62.7 percent. Single-stage adsorption at the pilot plant C_t = 30.4 would result in a loading of 7.0 percent and require 730 mg/l carbon. The computed two-stage countercurrent carbon requirement to deliver the same quality effluent would be 34 percent of 730 or 248 mg/l carbon. The pilot plant dosage of 145 mg/l was only 20 percent of single-stage or 58 percent of two-stage countercurrent adsorption system requirements.

This high-efficiency carbon utilization existed in the pilot plant process despite making the aforementioned allowance for the possible physical removal of filtrable suspended solids. Activated carbon, however, has been reported to have affinity for certain colloidal solids that produce haze or color and differences in performances of some carbons seemingly are related to their ability to remove a portion of the organic matter in suspension.

The pilot plant data supported earlier evidence (Figure 3) of an adsorption-resistant fraction of TOC and COD. A mean isotherm curve was prepared from the data of 35 individual composite-sample TOC isotherms (Figure 29). Adsorption proceeded reasonably similar to Freundlich theory up to carbon dosages in the range of 800 mg/l. At higher dosages TOC removals were less than theoretical. In Figure 29, adsorption at 800 mg/l carbon had removed 65 percent of the feed TOC. At 1600 mg/l removal went to 72.2 percent at a C_t of 6.5 mg/l whereas extending the curve in a straight line through the first three points would have resulted in about 79 percent removed at a C_t of 4.9 mg/l. Adsorbate removal would proceed in this range but at much reduced efficiency and Figure 30, a plot of applied TOC vs. TOC loading, shows this clearly. With the exception of three operating periods, all field results were with 145 mg/l carbon feed. The feed stream filtrable TOC ranged widely causing loading to range similarly however the removals persisted near a mean weighted value of 68.0 percent. The primary exceptions being the periods of higher or lower carbon dosage and the qualified alum study data. Higher dosages forced removal increases of only a few percentage points, thus indicating adsorptive capacity of the carbon was not the process limiting variable. Removals were limited to that fraction of the adsorbate indicators that could be adsorbed.

Filtrable COD removal by the pilot plant for all applicable periods averaged 69.9 percent (Figure 31). Considering pilot plant performance could not exceed the limit but rather approach it, it develops that carbon feeds other than about 130 percent of the feed stream filtrable COD are less than optimum. Three Operating Periods, 2, 4 and 15 produced mean COD/carbon loads of 53.6, 54.0 and 51.1 percent, respectively, placing the ultimate adsorptive capacity of the carbon in this range. Operating Period 15 was especially designed to avoid encountering the adsorbate limitation, whereas 2 and 4 were optimized coincidentally by the strength of the feed stream at those times. The adsorptive capacity must match the 70 percent of COD available for adsorption for optimum utilization, i.e., 100 x 0.70 COD/0.53 LOAD = 132% COD = mg/l carbon feed for the sewage studied. The isotherm in Figure 28 is from Period 2 with the carbon feed 127 percent of the feed filtrable COD for that date, making the isotherm evaluation consistent with the subsequent discussion.

Temperature - No regular program of temperature measurement existed. Random readings of the feed stream temperature ranged from a high of 91°F in July to a low of 75°F in February. A 1.5°F rise in temperature was recorded in the stream as it passed through the pilot plant on hot clear days. Although the adsorption process is temperature sensitive, the observed temperature range does not significantly affect the results and no adjustment in adsorption data was indicated.

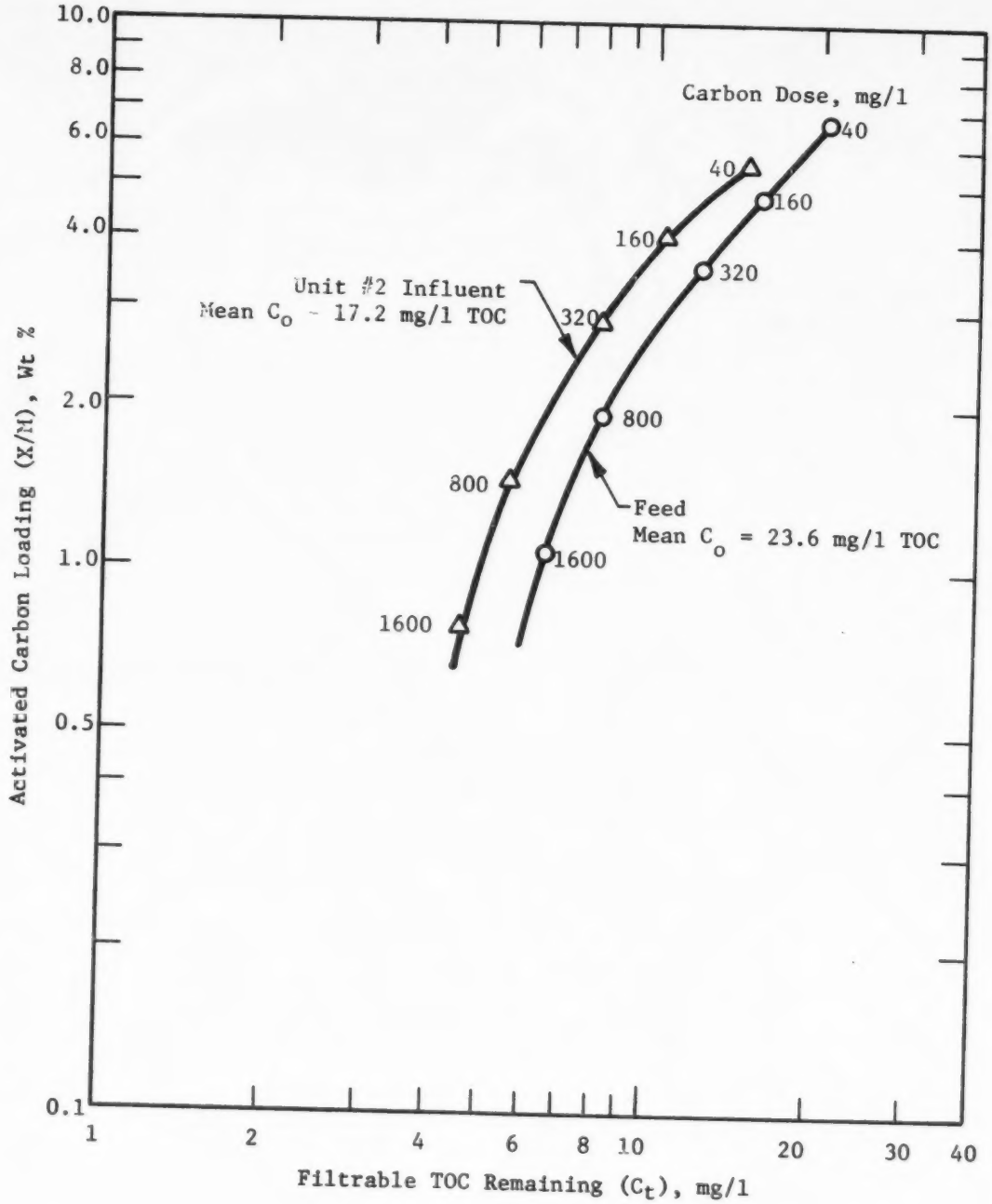

FIGURE 29: Mean Value Isotherms, TOC

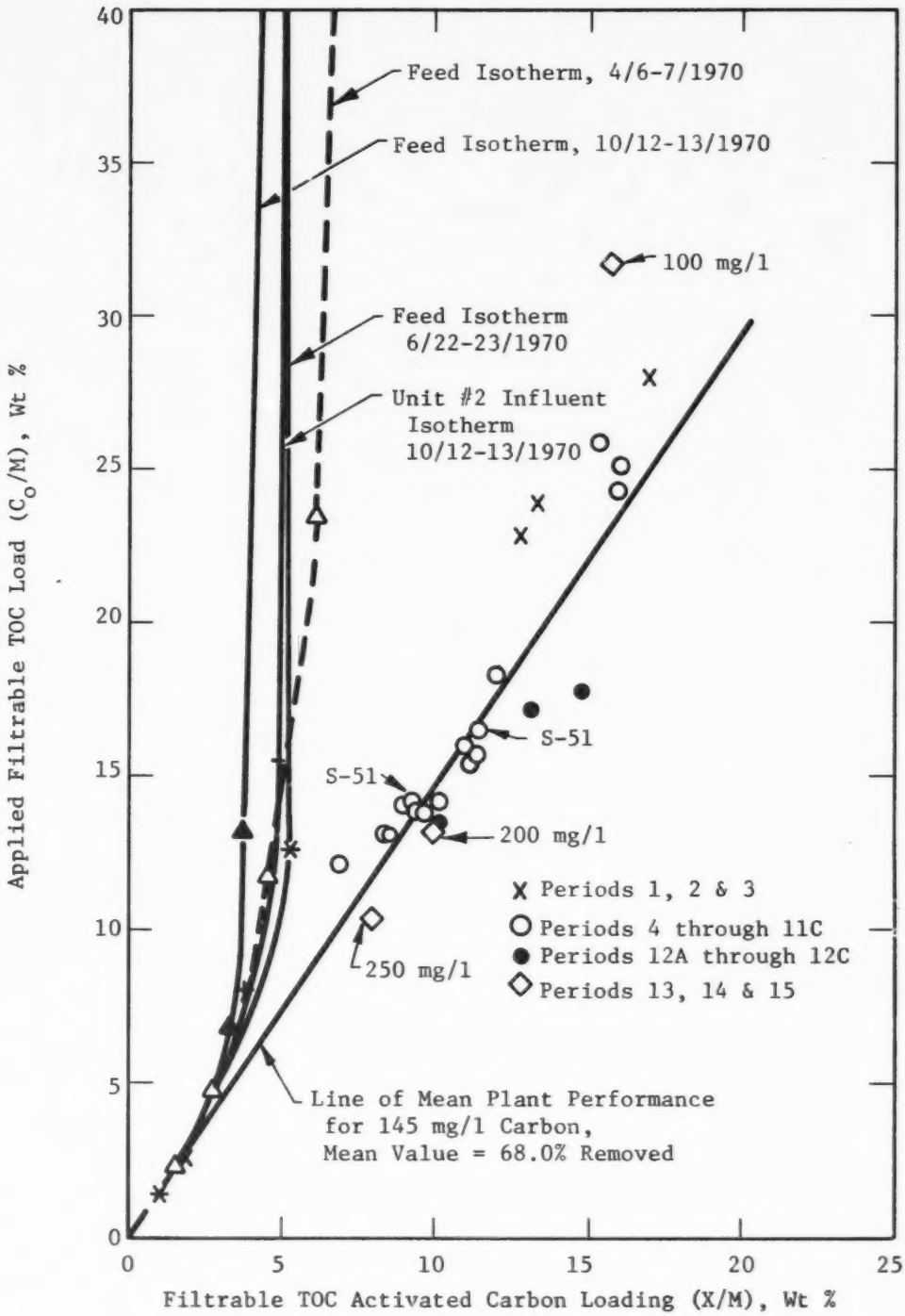

FIGURE 30: Clarifier Performance, Applied TOC Load vs. Carbon TOC Loading

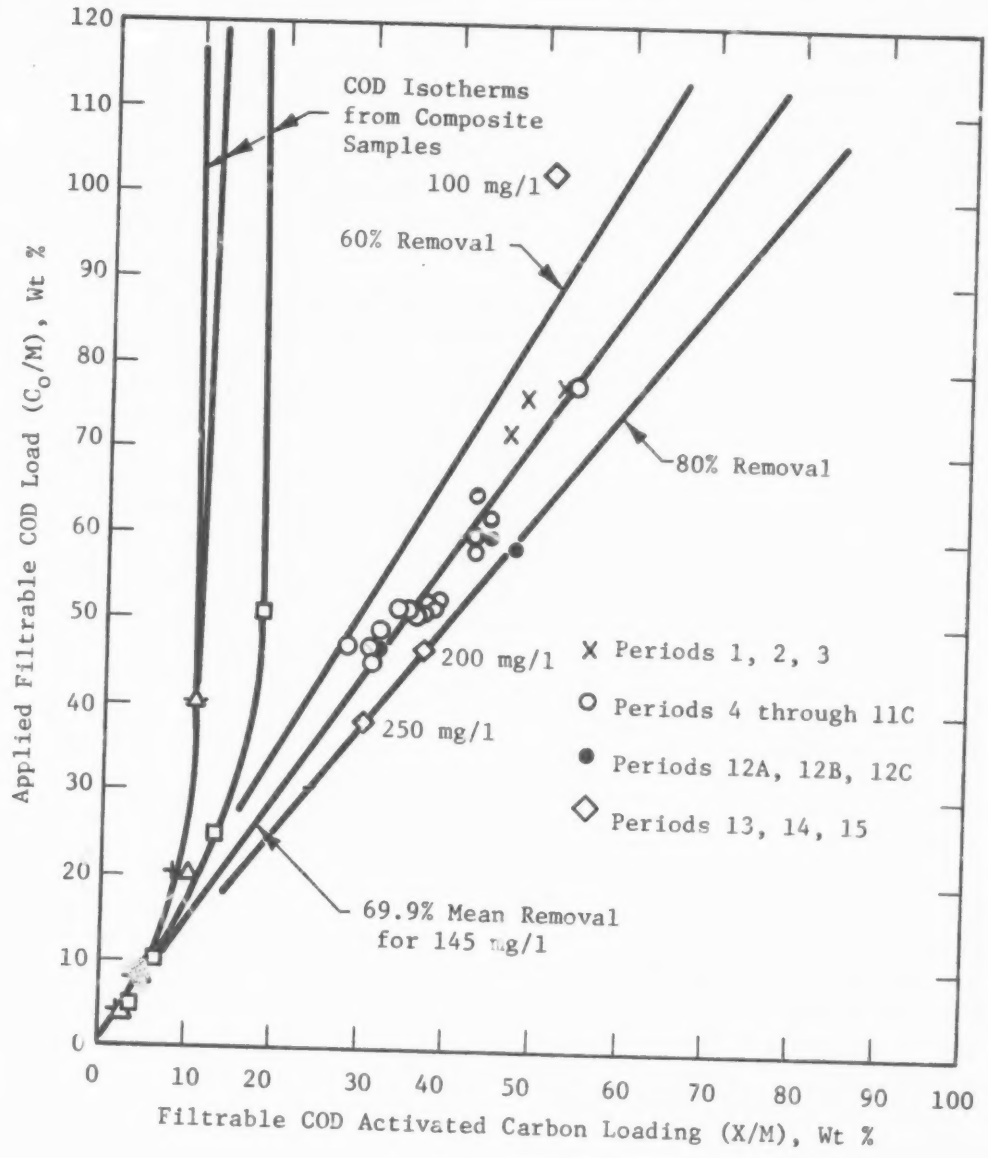

FIGURE 31: Clarifier Performance, Applied TOC Load vs. Carbon TOC Loading

pH - Composite samples were tested for pH during much of the program. The testing was halted after variations were found to be small. The feed stream pH ranged from 7.6 to 8.1, averaging 7.95. Unit #1 effluent ranged from 7.6 to 8.2, averaging 8.05. Unit #2 effluent ranged from 7.7 to 8.25, averaging 8.1. Gravity filter effluent ranged from 7.8 to 8.2, averaging 8.1. The sewage treatment plant effluent ranged from 7.4 to 8.1, averaging 7.9. The high alum dosages may have been responsible for lowering pilot plant effluent readings about 0.1 units during Phase 6.

Blowdown Suspended Solids - The suspended solids content of the slurry blowdown was determined on certain grab samples. The various configurations of the slurry concentrators and suspended solids influx rates generally resulted in unsuitable internal slurry concentrating conditions. During Operating Period 2 at a low overflow rate, however, conditions were suitable. Blowdown samples settled less than 5 Vol % upon 5-minute settling and were determined to contain 16.5 g/l and 24.6 g/l suspended solids for Unit #1 and Unit #2, respectively. Other data for Unit #1, adjusted by results of 5-minute settling tests made after collection, show 20 to 26 g/l at the higher carbon feed rates and 16 g/l at 100 mg/l carbon feed rate. Unit #2 averaged 22 g/l at the high feed rates dropping to 16 g/l for the 100 mg/l carbon test, possibly as the result of the high solids carryover noted earlier. Incomplete data were obtained from the Darco S-51 study but unadjusted results were 14 g/l for Unit #1 and 45 g/l for Unit #2. The suspended solids content of slurry to be regenerated influences regeneration costs.[12]

Color - The pilot plant process was substantially superior to the sewage plant process in terms of color removal. Figure 27 gives the mean color removal observed for both systems for comparable periods.

Nutrients - Nutrient removal was not an objective of this study but three sets of composite samples were analyzed for ammonia-nitrogen and phosphorus to assess any effect that might exist. The results on laboratory-filtered samples were as follows:

Sample Date	Feed	Unit #1 Effluent	Unit #2 Effluent	Filter Effluent	Sewage Plant
11/8-9/70					
Ammonia-Nitrogen as N	26	24	23	22	16
Phosphorus as P	10.0	9.1	7.8	5.8	9.9
11/9-10/70					
Ammonia-Nitrogen as N	25	23	23	21	11
Phosphorus as P	11.7	10.4	8.1	5.7	10.2
11/10-11/70					
Ammonia-Nitrogen as N	25	24	23	23	9.5
Phosphorus as P	10.9	8.8	4.6	4.6	10.6

An ammonia-nitrogen removal trend exists but not to a significant degree. Phosphorus removal, however, is significant. It is not known if this is associated with carbon adsorption, alum flocculation or phosphorus take-up by microorganisms.

Turbidity Measurement - Through most of the program, turbidity measurements were made on grab samples from the pilot plant at 2-hour intervals with a Hach Model 2100 Laboratory Turbidimeter because this unit has been used during other EPA powdered carbon treatment studies. Measurements were also made on the composite samples daily prior to the routine laboratory determinations with a Coleman #14 Spectrophotometer.

The Hach instrument measures scattered light and is standardized against a solidified Formazin suspension of known turbidity; hence its readings may be quite different from those obtained by light transmittance or absorption as in the spectrophotometer. The carbon content of the sample among other things affects the difference in readings between the two systems. Figure 32 compares the two readings for the composite samples. The feed stream ratio is the most consistent being devoid of activated carbon and other introduced variables. The clarifier effluents vary substantially as operation was varied. Unit #1 effluent was least affected of the two as a substantial portion of its turbidity was raw sewage suspended solids. Unit #2 effluent was highly variable as the sewage solids-to-carbon ratio was less. The filter ratio variation is not understood with the possible exception of Period 12 when carbon was observed penetrating the filter media. If the filter was discharging biological material produced therein, it would be unlike the sewage suspended solids or carbon solids monitored in the other samples.

Spectrophotometer determinations more closely approximated visual estimates of turbidity when activated carbon, either as an uncoagulated suspension or pinpoint floc, was present. This is substantiated by Figure 33. The ratios of spectrophotometer JTU to mg/l suspended solids vs. operating period curves center about the ratio of 1.0. The filter effluent curve again indicates a peculiarity exists in this sample stream.

The 2-hour field turbidity determinations provide the only directly measured information on diurnal feed stream variation and its effect upon the pilot plant. A 4-day period of these measurements is given in Figure 34. The cycling noted in the feed stream JTU is readily apparent in the clarifier effluents. Also, the inferred solids load on the pilot plant is relatively constant at a high level for 18 to 20 hours per day but drops off for 4 to 6 hours every morning. The curves also show the result of an operating accident. Loss of carbon feed for 8 hours had little effect on Unit #1 effluent but produced a measurable increase in Unit #2 effluent turbidity and loss of filtration efficiency for a number of hours.

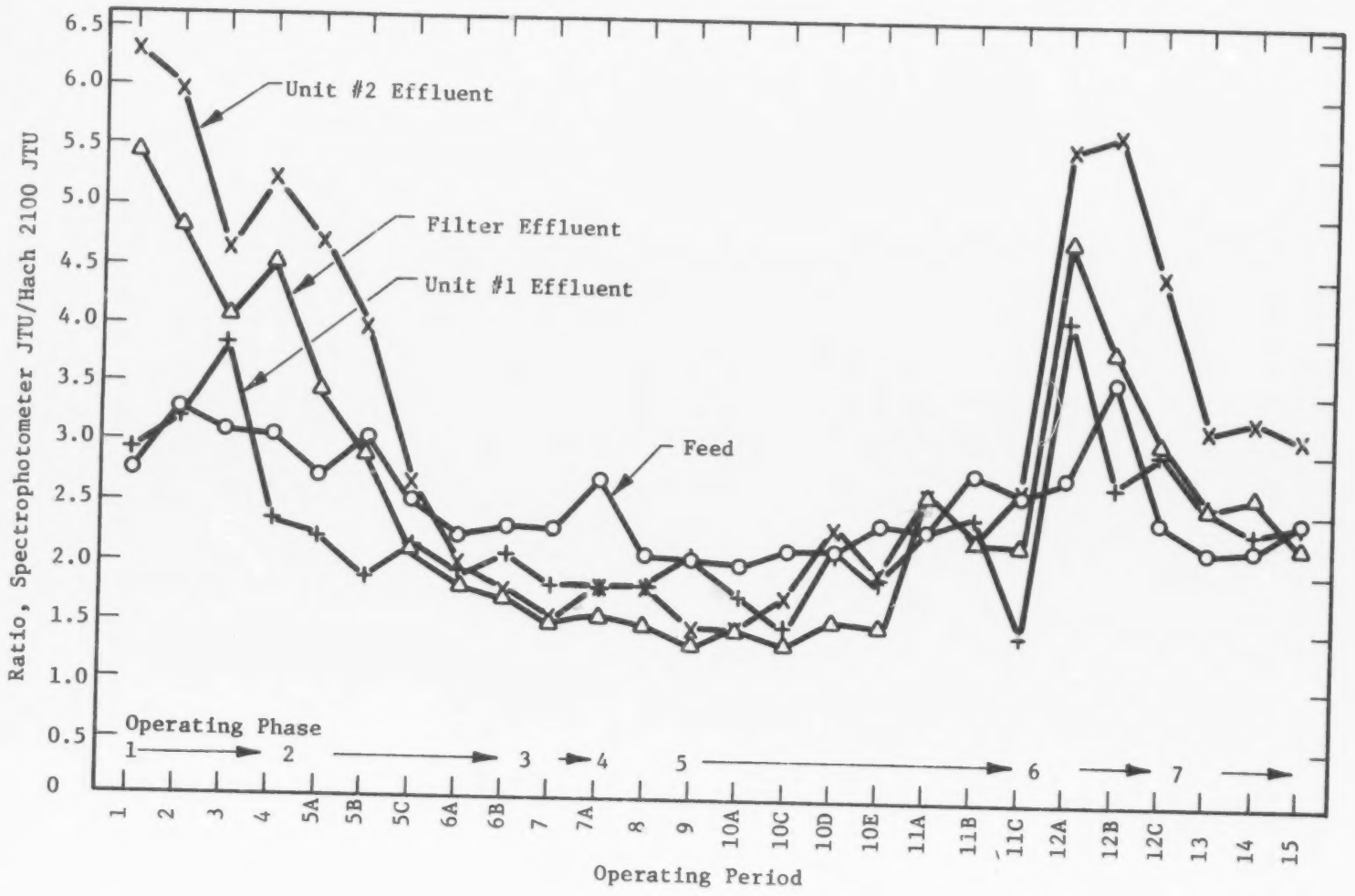

FIGURE 32: Ratio of Spectrophotometer JTU and Hach JTU

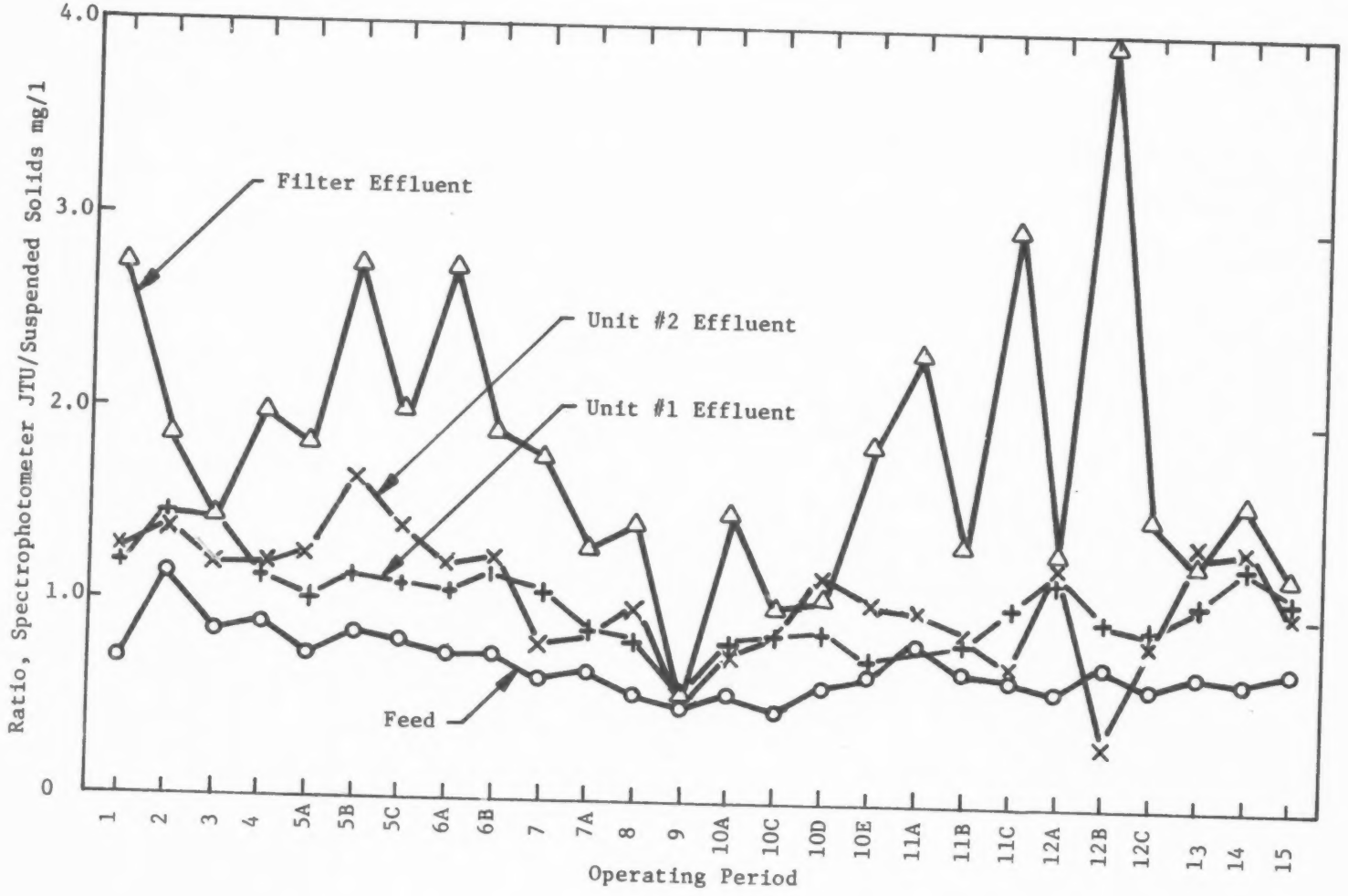

FIGURE 33: Ratio of Spectrophotometer Turbidity to Suspended Solids

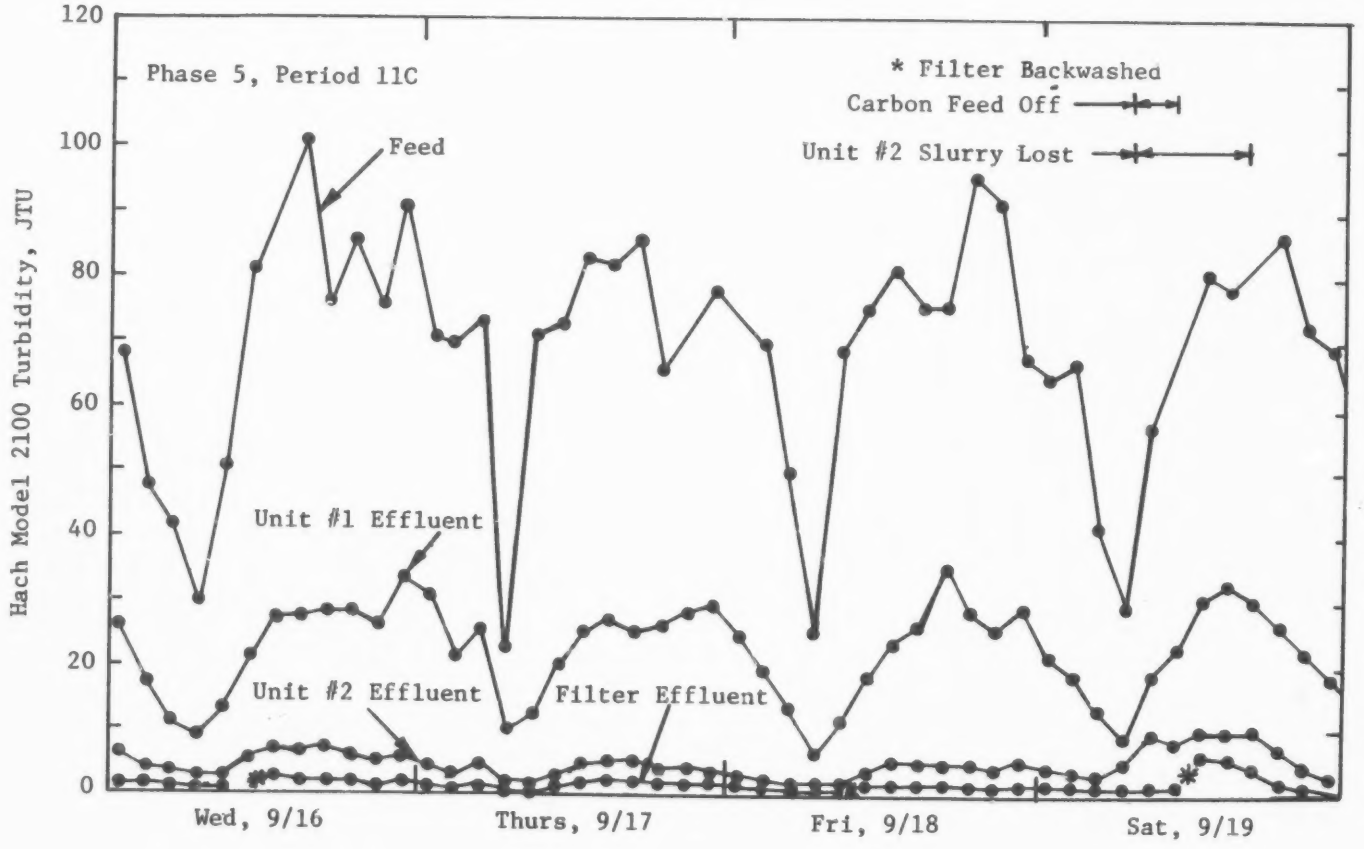

FIGURE 34: Diurnal Variation of JTU in System

Filter Operation - The duration of filter runs and the quality of filter effluent have been reported earlier in this report and are independent of the pilot plant operators. Wash water consumption, as a percentage of water filtered, has also been reported but this parameter was found to be highly dependent upon the pilot plant operators rather than design or process.

Backwashing procedure required establishing a low upflow rate for the duration of the surface raking followed by 5 minutes or more, as needed, of the high upflow wash rate. Without a flow-rate indicator the initial flow varied widely, as did the duration of the raking. The principal wash rate also varied with water temperature and the duration varied with the attention and judgment of the operator. Throughout the field program 12 individuals worked as operators at one interval or another, with the records showing water consumption ranging from a low of 325 gal. to a high of 1205 gal. per wash, both being extreme and unsatisfactory. Average consumption per wash by individual ranged from 481 to 809 gal. with the average for all backwashes being 647 gal. The median volume for all backwashes was 620 gal, however one experienced operator on a supervised shift averaged only 515 gal. with good results. Surface washing used 40 percent of this volume at a 10 gpm/ft^2 rate followed by the balance being consumed at 15 gpm/ft^2. Proper surface wash equipment would reduce the volume used prior to backwash to an insignificant figure; however, without the high rate manual surface wash period, the main backwash period would have to be extended with little change in the total water used. Based upon 515 gal. per wash, the wash water consumption would be in the range of 3.0 percent with filter surface loadings of 2.6 to 2.8 gpm/ft^2. This percentage is higher than would exist if a conventional filter, permitting head losses greater than the 4.5-ft limit of pilot plant equipment, were used.

DISCUSSION

General Considerations

Anthracite coal was selected for the top layer of filter media in the gravity filter to eliminate mud balls having a greater settling rate than the filter sand. In previous studies the polymer-flocculated carbon conglomerated surface filter sand into mud balls which settled to the top of the supporting gravel during backwashing. This deposited quantities of loaded carbon below the filtering zone, leading to progressive deterioration of effluent quality. Efforts to eliminate or control the problem were generally ineffectual. In the present study mud balls did not develop because of the surface raking which preceded each wash and any conglomerate would incorporate anthracite coal which would come to rest at the sand-coal interface effectively within reach of the surface raking. There was no danger of mechanically disturbing the supporting gravel which hindered the work in the previous study.

The final effluent was not completely satisfactory from an aesthetic viewpoint. It had a perceptible and distinctive sour odor (not hydrogen sulfide) quite disagreeable as compared to that of an activated sludge plant effluent. This odor could be detected in the clarifier effluents suggesting it was produced from some unstabilized and adsorption-resistant fractions present in the raw sewage. If it was present in the raw sewage it was masked by the stronger odors present prior to activated carbon contacting.

Since the materials contributing much of the TOC to the final effluent appeared to be adsorption resistant, some procedure other than direct carbon adsorption would be needed if further reduction in TOC was desired. The adsorption-resistant fraction probably varies between raw sewages and the concentration should be determined for prospective applications of the process. The raw sewage in one study[13] apparently contained less than 6 percent adsorption-resistant filtrable TOC permitting total TOC removals of 97 percent by adsorption. Tucson raw sewage with about 30 percent adsorption-resistant filtrable TOC was limited to about 90 percent total TOC removal. Both studies used the same carbon in evaluating the raw sewages. Earlier[6], the Tucson activated sludge plant effluent was successfully treated further by direct carbon adsorption indicating adsorption-resistant materials are either altered or assimilated by biological treatment.

Biological growth was evident on surface apparatus of both clarifiers and on the gravity filter sidewalls and underdrain system. The units were devoid of dissolved oxygen but slimes and facultative bacteria of various colors developed in the launders and on other interface apparatus. Green algae grew in the filter. Brushing loosened accumulations every few days. There was no observed manifestation of sub-

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surface biological activity in the clarifiers or in the slurry concentrators. There was no noticeable gas formation. Blowdown slurry odor was the same as noted previously for the effluent. During rare periods of inlet-screen bypassing or failure, raw solids settling with carbon floc on internal bracing in the clarification zone of Unit #1 would gas and pop to the surface. These clumps and other floatable material were manually skimmed away. Grease balls did not form in the system but some small ones entered with the influent. The scum baffle was very effective in eliminating floating material downstream. Flotation of carbon floc occurred in Unit #1 at times of carbon advance system malfunction. Entrainment of air at the carbon advance system surge tank due to accidental low liquid level approximated a dissolved air flotation system. The larger floatable solids were normally eliminated upstream of the pilot plant at the screened inlet system. A larger plant would, of necessity, incorporate a mechanical skimming apparatus.

Mean pilot plant performance in terms of TOC and COD removals are summarized in Table 8 along with comparable performance data for the activated sludge plant. Composite sample results for each of the powdered activated carbon dosages have been averaged without regard to coagulation system, slurry advance system or overflow rate. The data are probably biased in favor of the sewage plant. It was operated in a routine manner aside from some brief physical problem periods. In contrast, the pilot plant was purposely subjected to widely ranging experimental situations. Nevertheless, the pilot plant system out-performed the sewage plant most of the time.

Throughout the field program, carbon was fed into the returning slurry streams of each contactor-clarifier. The liquid phase of the slurry stream has the lowest adsorbate concentration in the unit. It is this liquid phase that becomes the unit effluent. The slurry advance system make-up liquid was taken from the clarification zone of Unit #1 to preserve this relationship. Countercurrent adsorption principles, i.e., always introduce the least loaded carbon into the most purified stream, were preserved in this study. Conversely, because the system was continuous, carbon seldom encountered feed stream adsorbate concentration because the feed stream was quickly diluted many-fold with recirculating slurry in the inner draft tube of the mechanism. This did not prevent loading the carbon to a high degree with adsorbate.

The total liquid volume of the pilot plant treating units was 1615 gal., excluding interconnecting piping. The retention time at 1.5 gpm/ft^2 overflow rate and 3 gpm/ft^2 filter rate would be 88 minutes from inlet chamber to filter effluent. An overflow rate of 2.25 gpm/ft^2 is considered a nominal maximum depending upon sewage/carbon characteristics at which the retention time would decrease to 63 minutes. At these overflow rates, the carbon contacting times would be 11.6 and 7.8 minutes, respectively, per contact unit. These times and overflow rates are more conservative than tested elsewhere.[13]

TABLE 8

Mean Treatment Plant Performance

Carbon Dosage mg/l	TOC				COD			
	Pilot Plant			Sewage Plant	Pilot Plant			Sewage Plant
	Inf mg/l	Eff mg/l	Red'n %	Red'n %	Inf mg/l	Eff mg/l	Red'n %	Red'n %
100 (ANA)	93.5	19.4	79.3	81.4	392.0	69.0	82.4	81.0
145 (ANA)	91.8	10.9	88.1	86.6	364.0	41.0	88.7	88.2
145 (S-51)	89.8	9.2	89.8	88.2	344.5	42.8	87.6	88.9
200 (ANA)	102.4	6.7	93.3	90.3	357.0	27.8	92.2	89.1
250 (ANA)	97.9	8.3	91.5	89.6	368.0	31.3	91.6	89.0

Samples not filtered in laboratory prior to analysis.

The solids-contact clarifier offers certain simplifications and economy for powdered activated carbon adsorption system design by combining in one vessel the functions of mixing, flocculation, and solids separation. Power use per stage is minimized and multiple tankage within each stage is eliminated, thus reducing space requirements and piping complexity.

The advantages of automated proportional chemical feed controls are not clear cut. For example, during the field studies there were occasions when carbon feed was nonexistent for up to 8 consecutive hours. On one occasion the carbon feeder was down 5.5 hours in the afternoon and the drive on Unit #2 was left turned off for 2 additional hours, all during the peak sewage strength period of the day. The composite samples for the 24-hour period showed the mean filtrable COD removal was 72.7 mg/l. The previous 24-hour period which was uneventful removed 67.6 mg/l. The feed streams tested 99.6 mg/l and 101.0 mg/l filtrable COD, respectively, while operating the plant at the maximum capacity of 1.5 gpm/ft^2 overflow rate. The carbon feed was off for nearly 9 times the single clarifier retention time. Originally, data for days with underfeed were automatically discarded but later, because the effect could not be distinguished from the normal daily deviation, these data were retained.

The estimated 7 to 9 lb of carbon inventory in the recirculating slurry had adequate adsorption capacity to span long periods without feed. Difficulties were associated with maintaining slurry inventory, effluent clarity and floc settleability but not adsorption. If carbon feed need not be varied as the sewage strength changes then coagulant feeds may not be proportioned to sewage strength either as carbon flocculation must be maintained. Proportional feed based upon flow rate for a variable rate plant is, however, highly practical and would result in chemical economy as compared to feeding the minimum marginal dosages for peak strength and flow rates continuously.

The pilot plant was a constant rate system so no adjustment in the chemical consumption data is justified for flow rate proportioning.

Periods of weak feed sewage were 4 to 5 hours per day or 17 to 21 percent of time; however, chemical feeds were marginal for peak feed strength periods so proportioning by feed stream turbidity measurement would simply shift consumption from weak to strong periods. Again, no adjustment in overall chemical economy is thought justified.

Regeneration and reuse of powdered activated carbon combined with raw sewage solids and alum content is presumed possible. Development of various processes is proceeding. At least one investigation[13] shows promise, although the ratio of carbon to raw sewage solids was much higher than in this study. Alum recovery and reuse were also considered in that investigation.

Process Economics

Based upon the design approach noted below, cost estimates have been developed for 10-mgd and 100-mgd treatment systems of the type studied.

Plant Design - Each plant incorporates complete water recovery, eliminating any requirement for separate disposal. All process water needs are met with settled filter wash water which is pumped from storage to fill slurry makeup tanks and to dilute polymer in day tanks. Excess backwash water is pumped to the downstream adsorption units for recovery.

Chemical feed systems are sized to provide up to 300 mg/l, 30 mg/l and 100 mg/l of carbon, polyelectrolyte and alum, respectively. Carbon recovery via regeneration is visualized and 15 percent loss per cycle is assumed.

The site requirement for the 10-mgd plant is an area 110 ft x 150 ft. Two ACCELATOR mechanisms in basins 62 ft square x 17.5 ft deep handle the adsorption duties of this plant and three 4-cell gravity sand-anthracite filters with GREENLEAF ® Filter Controls (a unique siphon valve system for controlling the functions of multiple rapid-sand gravity filters) provide effluent polishing at 3 gpm/ft^2. These concrete structures, as well as a two-story metal building housing the carbon regeneration system, polyelectrolyte feed tanks and pumps, alum feed tanks and pumps, carbon makeup and polyelectrolyte storage, office and laboratory are constructed above grade.

Agitated covered concrete tanks located below grade between the ACCELATOR basins and filters include three 4000-ft^3 carbon-slurry day tanks. During operation, one tank is in use, a second is being readied with water plus carbon from regeneration and/or storage, and the third is ready and on standby for use the following day. Two 6000-ft^3 mixed tanks provide up to 5.4 hr of holding time for blowdown from the respective clarifiers, one supplying partially spent carbon to the first stage unit and the other furnishing fully spent carbon to regeneration.

The type of filter visualized requires no wash water storage. A 21,000-ft^3 unmixed basin for backwash wastewater catchment completes the below-grade tankage for this plant.

At a feed rate of 140 mg/l (1.17 lb/1000 gal.), the daily carbon requirement is 11,700 pounds. The plant requires an in-service carbon inventory of 31,300 lb and storage for one carload of bagged carbon. A 20-ton carload will meet maximum makeup requirements for 23 days. Drum delivery of polyelectrolyte is assumed. Consumption at 14 mg/l is about 16 drums per week. At a feed rate of 30 mg/l, liquid alum (50% alum dry wt.) would be consumed at 5,000 lb per day. A tank car load would last 16 days.

A 100-mgd plant of similar general design occupies an area 450 ft x 450 ft. To provide appropriate flexibility and emergency operating capability, five 20-mgd ACCELATOR units operate in parallel as the upstream contact stage, with five similar parallel units as the downstream stage. These mechanisms are installed in common-wall concrete basins, each 88 ft square x 21 ft deep. Fourteen 43-ft square, 4-cell filters with GREENLEAF Filter Controls are required.

A two-story building covers two below-grade agitated blowdown and three carbon slurry day tanks and houses the carbon regeneration system, two polyelectrolyte day tanks, carbon feed systems, alum feed systems, service pumps, offices and laboratory. Unit capacities of the blowdown and carbon slurry tanks are 48,000 ft^3 and 40,000 ft^3, respectively.

In this plant, a conventional 67-ft diameter clarifier is used for filter wash water recovery. The overflow furnishes all plant process water and underflow solids are transferred by gravity to the first stage blowdown tank enroute to regeneration. As before, excess filter backwash wastewater goes to the downstream clarifiers for recovery.

The in-service carbon inventory for this plant is about 300,000 pounds. At 140 mg/1, daily carbon use is 117,000 lb and a make-up requirement of 17,550 lb/day is indicated. Bulk storage is not included. Rail delivery of bulk carbon is visualized with cars unloading make-up quantities into day tanks as required.

Polyelectrolyte coagulant is delivered by tank car to a 20,000-gal. storage tank from where it is pumped as required to either of two day tanks for dilution to feed strength. The dilute material is then pumped to each contact-clarifier through variable-orifice rate control devices installed along a distribution header serving all units.

Liquid alum would be pumped directly to the application points from tank cars through a similar distribution system.

Both plant designs assume a mild climate location with weatherproof equipment installed in the open. Carbon handling equipment is either of corrosion resistant material or is protected with suitable coatings. Alum handling equipment is similarly constructed. Essential instrumentation for flow metering and control and low-lift process and service pumps are included. A conventional intake, comminutor and grit removal structure precedes the low lift pumps.

Treatment Cost - The estimated construction cost of the 10-mgd plant, including land but exclusive of the capital cost of a carbon regeneration system, is as follows:

Equipment delivered ($295,000) and installed	$366,500
Plant structures	236,700
TOTAL PLANT COST	$603,200
Engineering (10% of total plant cost)	60,300
Contractor's fee (10% of total plant cost)	60,300
Land acquisition (2% of total plant cost)	12,100
Contingencies and omissions (15% of total plant cost)	90,500
TOTAL CAPITAL COST	$826,400

Annual operating cost breaks down as follows:

Capital ($826,400 for 20 years @ 6%)	$ 72,000
Maintenance (3% of equipment + 1% of structures)	13,400
Labor	79,900
Power (1¢/KWH)	21,400
TOTAL	$186,700

This is equivalent to 5.1¢/1000 gal. treated, exclusive of chemical costs. The latter are influenced by the feed-stream quality and treatment objectives. For this study the feed stream is assumed similar to that encountered during the field period and optimum carbon dosage of 132% of mean filtrable COD is used. Mean filtrable feed COD was 89.3 mg/l for all composite samples analyzed so the carbon feed is 118 mg/l. A polymer feed of 14 mg/l and alum feed of 25 mg/l are the estimated quantities to control the process.

Carbon regeneration will supply 85 percent of the carbon feed with the residual 15 percent being new carbon makeup. Carbon regeneration costs of Berg, et al.,[12] were modified to reflect 15 percent solids rather than 25 percent solids feed to the thermal reactor. Centrifugation costs of 0.2¢/lb are included in the regenerated-carbon unit cost. Estimated chemical costs are:

	¢/1000 gal.
Regenerated carbon (0.983 lb/1000 gal.) (1.9¢/lb) (0.85)	1.6
Makeup carbon (0.983 lb/1000 gal.) (9.5¢/lb + 3¢/lb freight) (0.15)	1.8
C-31 polyelectrolyte (0.117 lb/1000 gal.) (31¢/lb + 3¢/lb freight)	4.0
Liquid alum (0.208 lb/1000 gal.) (3.6¢/lb + 3¢/lb freight)	1.4
TOTAL	8.8

On this basis, the total cost of plant operation on this scale is 13.9¢/1000 gal.

Using the same approach, the estimated construction cost of the 100-mgd installation is as follows:

Equipment delivered ($2,120,400) and installed	$2,374,600
Plant structures	1,270,600
TOTAL PLANT COST	$3,645,200
Engineering (10% of total plant cost)	364,500
Contractor's fee (10% of total plant cost)	364,500
Land acquisition (2% of total plant cost)	72,900
Contingencies and omissions (15% of total plant cost)	546,800
TOTAL CAPITAL COST	$4,993,900

The breakdown for annual operating cost exclusive of chemical cost is:

Capital ($4,993,900 for 20 years @ 6%)	$ 435,400
Maintenance (3% of equipment + 1% of structures)	83,900
Labor	239,800
Power (1¢/KWH)	122,500
TOTAL	$ 881,600

This is equivalent to 2.4¢/1000 gal. treated. Of the unit chemical costs noted previously, only that for regenerated carbon might decrease because of larger-scale facilities. The others would remain unchanged. Using the original chemical cost figures, the total cost of treatment in the 100-mgd plant would be 11.2¢/1000 gal.

Unlike some applications which might have a wide ranging carbon feed requirement, treatment of sewage at any one location would be characterized by a moderate feed range, say plus or minus 30 percent of the example dosage. Treating costs would vary plus or minus 1¢/1000 gal. for optimized conditions. The estimated 1.9¢/lb for carbon regeneration cost may be quite speculative in view of the raw sewage solids content however it is based upon a conservative 15 percent solids delivered by a centrifuge. The unit cost may go as low as 1.5¢/lb for 25 percent solids or as high as 5.0¢/lb for uncentrifuged 5 percent solids in the slurry. The former would lower the estimated total cost of the carbon feed to 3.1¢/1000 gal. and the latter would raise it to 6.0¢/1000 gal. as compared to 3.4¢/1000 gal. in the tabulation.

A 10-mgd high-rate powdered activated carbon treatment plant[13] not utilizing countercurrent contacting was estimated to cost 21.5¢/1000 gal. Carbon regeneration was 1.5¢/lb; however, the single-stage process required 600 mg/l carbon feed with high alum feeds and polymer to control same. Chemical operating costs totaled 17.4¢/1000 gal. with alum recovery and reuse. With the greater carbon feed rate, i.e., 5 lb/1000 gal., cost of regeneration was 6.75¢/1000 gal. and makeup carbon cost 4.5¢/1000 gal. for a total of 11.25¢/1000 gal. This approximates the cost of using virgin carbon, freight included, on a throw-away basis in the two-unit countercurrent process.

The projected total cost, some years ago, of a 10-mgd granular activated carbon adsorption system using multiple downflow contact columns was 8.3¢/1000 gal.[4] A plant of this type might require as much as 133,000 lb of in-service carbon inventory for each 1000 gpm of system capacity. The powdered activated carbon system described is conservatively estimated to require only 4,500 lb of in-service carbon inventory for each 1000 gpm of capacity. These inventories represent an initial carbon investment of $240,000 (26¢/lb) for a 10-mgd granular carbon plant as compared to only $3,100 (9.5¢/lb) for a powdered carbon plant. Both systems require facilities to warehouse and handle 10 - 20 percent makeup carbon.

At 50 to 55 percent carbon loading, powdered carbon utilization would be about 80 to 100 percent of the top utilization reported for granular systems depending upon the extent of biological degradation assisting the granular system. More frequent regeneration might be required for the powdered carbon, but this factor is more than offset by the normal price differential between powdered and granular carbons. There would be problems applying raw sewage to carbon columns.

The present cost of activated sludge treatment is estimated to be 17¢/1000 gal. for a 10-mgd plant and 8.2¢/1000 gal. for a 100-mgd plant, 1967 cost information[14] was adjusted to 1971 by applying the EPA construction cost index and updated amortization (20 years @ 6%) to plant cost and 5 percent per year increase in operating and maintenance costs. The pilot plant process is less costly than the activated sludge process at the 10-mgd plant size. As unit chemical costs do not diminish greatly with increasing plant size the activated sludge process becomes less expensive for large plants.

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APPENDIX

Low-Level COD Determination

Discussion - The procedure is fundamentally that given for dilute samples in "Standard Methods for the Examination of Water and Wastewater," 12th Edition, American Public Health Association, p. 513 (1965), amended to further minimize the introduction of extraneous-material error.

The published precaution concerning contaminated glassware and atmosphere is reemphasized. In addition, laboratory grade distilled water or demineralized water used for reagent preparation, pre-reflux dilution, blanks, rinsing, or final dilution may cause erratic results. Water used for these functions should be especially prepared by the distillation procedure described fully in the "Reagents" section for blank water.

Contamination of the sulfuric acid reagent, even though carefully prepared, can introduce an error. The error is cancelled by running all samples and blanks to a set with the same batch of Ag_2SO_4-H_2SO_4 solution.

No significant error has been attached to the use of a polyethylene wash bottle for rinsing down condenser tubes and tips with the blank water.

The apparatus described is effective in preventing dust contamination and in protecting the analyst from injury by bumping which may occasionally occur during refluxing. It differs slightly from that described in "Standard Methods."

Apparatus -

500-ml short-necked round-bottom boiling flasks. ST joint 24/40, Corning #4320 or equal.

Allihn type condensers, 500-ml jacket, drop tip inner ST joint 24/40, top outer ST joint 24/40, Corning #2480 or equal.

Connecting tubes 75°, both ends inner ST joints 24/40, Corning #8920 or equal.

Tube stoppers, cap type, full length outer ST joint 24/40, 2 required per condenser.

Heaters, Precision Catalog #61560 or equal.

Running-water cooling bath large enough for the required number of flasks.

The apparatus is mounted such that the complete glass assembly consisting of flask, condenser, and upper connecting tube may be raised from or lowered to the heater as required. The connecting tube is placed on the upper end of the condenser pointed away from work or chemical storage areas. The cap-type stoppers are placed on the condenser tip and connecting-tube tip to prevent entrance of dust particles when the apparatus is not in use. Both caps are removed when the apparatus is in use.

Reagents -

Blank water: Redistill a quality distilled or demineralized water that has been passed through a bed of activated carbon (from which most of the chloride has been leached) in the following manner.

Place in a 3-liter round bottom distilling flask containing several boiling chips or beads, 500 ml pretreated water, 200 ml 0.25N (approximate) potassium dichromate, 200 ml sulfuric acid reagent and about 1 g $HgSO_4$. Swirl to mix and dissolve the $HgSO_4$. Mark the level of this mixture on flask. Add about 1.5 liter of pretreated water and mix. Bring to a boil on the electric heater. If a short-necked ST flask is used, a long glass stirring rod may be kept in the flask while mixture is heating to prevent bumping. Stir occasionally until boiling starts. Allow to boil for 10 minutes wasting steam to the atmosphere. Place the connecting tube and condenser, and waste steam through the uncooled condenser for 2 or 3 minutes. Start the cooling water and waste enough condensate to rinse condenser.

Collect the distillate in a glass bottle rinsed several times with distillate. A 9-lb acid bottle, well cleaned with chromic acid, makes a good container. Use an adapter from the condenser tip into bottle. The adapter tip should enter the receiver bottle through a hole in a loose fitting foil dust cap. Distill until the level in the flask drops to the mark.

Cap the water storage bottles when not in use.

The oxidizing mixture in the flask may be used repeatedly until there is obvious discoloration.

Standard potassium dichromate solution, 0.025N: Prepare per "Standard Methods" except prepare with blank water.

Sulfuric acid reagent: Place 21 g reagent grade silver sulfate in a newly opened standard 9-lb bottle of concentrated reagent-grade H_2SO_4. With a flat tipped large diameter glass stirring rod, carefully crush any lumps of Ag_2SO_4 and stir thoroughly, repeating as often as necessary. Complete dissolution may be obtained in 20 minutes or less. Cap securely and mix thoroughly by inversion.

Standard ferrous ammonium sulfate titrant, 0.01N: Prepare per "Standard Methods," or preferably, 0.025N solution may be substituted and has been found to give reproducible results.

Ferroin indicator solution: Prepare per "Standard Methods."

Silver sulfate, reagent powder.

Mercuric sulfate, analytical grade crystals.

Procedure -

Set up two blanks for each set of samples.

Place 5 ml concentrated H_2SO_4 in identified 500-ml boiling flasks containing about 7 glass beeds. Keep flasks capped with small glass beakers at all times except when introducing samples and reagents.

Add approximately 0.5 g $HgSO_4$ and swirl to mix until $HgSO_4$ is dissolved.

Add 50.0-ml blanks and samples, or aliquots diluted to 50.0 ml with blank water.

Add 25.0 ml 0.025N potassium dichromate solution swirling to mix during addition.

Remove caps from reflux condensers and top connecting tubes. Start the cooling water through the condenser jackets. Turn heaters on and set at high heat.

To each flask, add without mixing 70 ml Sulfuric acid reagent (Ag_2SO_4-H_2SO_4 solution).

Taking each flask in turn swirl to mix contents quickly but thoroughly and connect immediately to the condenser. Clamp and lower the assembly to the heater. Boiling should start quickly without bumping. When vigorous boiling is established, turn the heater down to a setting to maintain boiling. Reflux for 2 hours.

At the end of the reflux time, taking each unit in turn, in the same order as before, turn off the heater and raise the apparatus several inches above the heater. Place an asbestos mat over the heater. Allow to cool for 15 minutes.

Remove the connecting tube and rinse down the condenser with a little blank water. Allow about 10 - 15 seconds for draining then lower the flask from the condenser. Rinse the condenser tip into the flask with a little blank water and cap. Place the capped flask in the cooling bath immediately. Replace the condenser and connecting tube caps.

While the flasks are cooling, prepare a ferrous ammonium sulfate standardization. Place 25.0 ml of 0.025N potassium dichromate solution and about 200 ml blank water in a 500-ml Erlenmeyer flask. Add with swirling, 30 - 35 ml concentrated H_2SO_4. Cover and place in the water bath to cool. NOTE: If using 0.01N ferrous ammonium sulfate, use 10.0 ml of 0.025 potassium dichromate solution for the standard.

When the reflux flasks have cooled (cool to the touch), carefully add 200 ml blank water to each. Mix and return them to water bath.

When all flasks, including the standard, have cooled, add 2 - 3 drops of ferroin indicator to each. Titrate with ferrous ammonium sulfate directly in the boiling flask. Do not transfer. This is essential for consistent results.

Empty the flasks without removing the beads. Rinse the flasks and beads thoroughly 3 times with distilled water and 3 times with small portions (about 10 ml) of blank water. Replace the caps immediately.

General Notes - The blanks should titrate to within 0.2 ml of the standard if there is no contamination. Different lots of Ag_2SO_4-H_2SO_4 solution will produce different blank titrations. With careful attention to cleanliness there will be less than 0.2 ml difference between blanks. Within this limit use the higher blank titration result, not the average. For very precise work (non-routine) the two blank titrations should agree.

All glassware must be kept clean and dust-free. Use cleaning acid followed by distilled water and finally blank water. To illustrate the relevance of contamination, calculations indicate that in a 50-ml sample cellulose dust weighing 0.05 mg would contribute about 1 mg/l COD.

Calculations - Calculate mg/l COD per "Standard Methods."

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